Tiina Koskela

Removal of hardness from groundwater with nanofiltration

Case study: Meri-Lapin Vesi Oy

Helsinki Metropolia University of Applied Sciences

Bachelor of Engineering

Environmental Engineering

Thesis

16.08.2016

Author Title Number of Pages Date	Tiina Koskela Removal of hardness from groundwater with nanofiltration Case study: Meri-Lapin Vesi Oy 51 pages + 5 appendices 16.08.2016
Degree	Bachelor of Engineering
Degree Programme	Environmental Engineering
Specialisation option	Waste and Water Treatment Technology
Instructors	Leena Vänskä, Process Designer, Pöyry Finland Oy Esa Toukoniitty, Senior Lecturer, Metropolia UAS

This thesis was done for Meri-Lapin Vesi Oy. It studied the removal of hardness from groundwater with nanofiltration. The thesis continued the testing of nanofiltration that was already done a few years ago in Meri-Lapin Vesi. The focus of the thesis was on evaluating the operational costs of the nanofiltration in Meri-Lapin Vesi case and on eliminating the error factors of the previous study.

The new tests were done in Meri-Lapin Vesi facilities in spring 2016. DOW FilmTech nanofiltration membrane NF90-4040 was tested in two different cases for groundwater. The pilot scale filtration rig was added to the end of the current treatment process to see how it removes hardness.

The feed water contains a large number of hardness ions: calcium and magnesium. Also iron and manganese are still found in the feed water. The current process in the plant for the ground water is an iron and manganese removal process. The nanomembranes had a hardness removal rate over 99 %. A recovery rate of approximately 72 % was obtained at Meri-Lapin Vesi condition when a recovery rate of 80 % was aimed at. The system was driven at a flux rate of 30 – 60 l/m^2h.

Some difficulties with the water quality that Meri-Lapin Vesi has were encountered. For example the fouling of the membranes was noted to be a problem. This was assumed to be due to the organic matter since the feed water permanganate value was at the upper end of the quality recommendation. Otherwise, the nanofiltration suits well for softening the water Meri-Lapin Vesi uses.

An operational cost evaluation was made for the filtration system that was used in the tests. Calculations were made based on the results of the tests. The estimated price fitted in the price range of other studies that are made around the world. Nevertheless, operational costs in the other studies are not fully comparable with each other or to this study. This is due to the fact that the studies have been conducted in a large time span in several different countries.

Keywords	nanofiltration, hardness, hardness removal, softening

Tekijä Otsikko Sivumäärä Päivämäärä	Tiina Koskela Removal of hardness from groundwater with nanofiltration Case study: Meri-Lapin Vesi Oy 51 sivua + 5 liitettä 16.08.2016
Tutkinto	Insinööri (AMK)
Koulutusohjelma	Ympäristötekniikka
Erikoistuminen	Jätteen- ja vedenkäsittelytekniikka
Ohjaajat	Leena Vänskä, prosessisuunnittelija, Pöyry Finland Oy Esa Toukoniitty, lehtori, Metropolia AMK

Tämä opinnäytetyö tehtiin Meri-Lapin Vesi Oy:lle. Opinnäytetyössä tutkittiin nanosuodatuksen soveltuvuutta kovuuden poistoon pohjavedestä. Opinnäytetyö on jatkumoa muutama vuosi aikaisemmin suoritettuihin testeihin, joissa Meri-Lapin Vesi koeajoi nanosuodatusta. Työssä keskityttiin tutkimaan nanosuodatuksen käyttökustannuksia Meri-Lapin Veden tapauksessa. Tavoitteena oli saada eliminoitua edellisten kokeiden virhe tekijät, jotta käyttökustannukset nanosuodatukselle saadaan laskettua.

Uudet kokeet tehtiin Meri-Lapin Veden tiloissa keväällä 2016. DOW FilmTechin NF90-4040-nanosuodatuskalvoja koeajettiin kahdella eri laatuisella pohjavedellä. Pienen mittakaavan nanosuodatuslaitteisto lisättiin koeajon ajaksi laitokselle. Laitteisto sijoitettiin nykyisen käsittelyprosessin jälkeen, testaamaan kovuuden poistoa.

Syöttövedessä on paljon kovuutta aiheuttavia kalsium- ja magnesiumioneita. Myös rautaa ja mangaania on havaittavissa syöttövedessä. Laitoksen nykyinen prosessi pohjaveden käsittelyssä on raudan ja mangaanin poistoprosessi. Nanosuodatus poisti yli 99 % kovuusioneista. Koeajoissa tavoiteltiin 80 %:n saantoa, mutta koeajojen saanto jäi keskiarvoltaan noin 72 %:iin. Suodatusta ajettiin noin 30 – 60 l/m²h vuolla.

Koeajojen aikana todettiin muutamia hankaluuksia syöttöveden kanssa. Suodatuskalvojen tukkeutuminen osoittautui ongelmalliseksi. Tämän oletettiin johtuvan vedessä olevasta orgaanisesta aineesta, sillä syöttöveden permanganaattiluku oli laatuvaatimuksien ylärajalla. Muuten nanosuodatus toimi hyvin Meri-Lapin Veden olosuhteissa ja poisti hyvin kovuutta.

Käyttökustannustarkastelu tehtiin suodatusjärjestelmälle, jota käytettiin koeajoissa. Laskelmat tehtiin koeajon tuloksien pohjalta. Laskelmien perusteella saadut käyttökustannukset sopivat hintahaarukkaan, joka perustuu muista nanosuodatustutkielmista saatuihin tuloksiin. Tämän tutkimuksen ja muiden tutkimuksien käyttökustannukset eivät kuitenkaan ole suoraan verrannollisia keskenään. Tämä johtuu siitä, että tutkimukset ovat tehty erittäin laajalla aikavälillä ja ympäri maailmaa. Käyttökustannusten vertailu on enemmänkin suuntaa-antava.

Avainsanat	nanosuodatus, kovuus, kovuuden poisto, pehmennys

Contents

1 Introduction

Meri-Lapin Vesi Oy is a joint public limited company supplying water in Kemi, Tornio, Keminmaa and Tervola. Meri-Lapin Vesi Oy was founded in 1997 with a mission to provide good quality water. Several water intakes ensure enough water for the clients' needs. Meri-Lapin Vesi supplies treated groundwater for their clients.

Meri-Lapin Vesi has several groundwater reservoirs. Some of the reservoirs have high water hardness. The company may have to use these reservoirs more in the future to ensure the increasing demand of water. The supplied treated water fulfils The Ministry of Social Affairs and Health's requirements for drinking water. Water hardness is not defined in the ministry's quality requirements; nevertheless, the company wishes to lower the hardness of the water.

Meri-Lapin Vesi Oy has previously commissioned a study to research the removal of hardness from groundwater. In this research nanofiltration was mentioned as an option for further study. Also in 2013, Meri-Lapin Vesi made pilot tests with nanofiltration to see if it suits for their need. In this small-scale test run, it was found out that nanofiltration is suitable for removing hardness from the groundwater.

Nanofiltration is a water purification method where particles from water are filtrated through the membrane. Nanofiltration was developed in 1970 and became more common in late 20th early 21st century. Several studies on the particle removal properties and operational costs of nanofiltration have been conducted around the world.

In this thesis, new test runs were made with the most promising membrane from the previous test. The main reason for conducting these new test runs was Meri-Lapin Vesi's interest in determining the operational costs of nanofiltration. In the previous test operational cost was also calculated but the aim in this new study was to eliminate possible error factors from the previous study. On the basis of the new test runs, the operational costs can be recalculated and adjusted.

2 Literature review

Literature review focuses on explaining water hardness and nanofiltration. A huge number of studies on nanofiltration have been conducted around the world. This chapter summarises the findings of the previous studies and examines what kind of operational costs were estimated in them.

2.1 Water hardness

Hardness (Hr) is a feature in a water quality that causes problems for water consumers. Water hardness is caused by multivalent cations such as calcium (Ca^{2+}) and magnesium (Mg^{2+}). They are the most common ions causing the hardness in water and other cations are part of the cause too but often neglected. Hardness is defined as a total sum of cations in the water. [1,2] Total hardness can be calculated with the following formula:

$$Hr = C(Ca^{2+}) + C(Mg^{2+})\ [\mathrm{mg/l}]$$

If a numeric value for calcium concentration in the water is known, the hardness can be expressed as calcium carbonate in the water. This can be done because the molar masses of the calcium and calcium carbonate are known. When the ratio between calcium carbonate is known, it can be used as a multiplier for the concentration of the calcium. The same holds true for magnesium when its concentration is known. [3]

$$\frac{M_{\mathrm{CaCo_3}}}{M_{\mathrm{Ca}}} = \frac{100.1\ [\mathrm{g/mol}]}{40.1\ [\mathrm{g/mol}]} = 2.5$$

$$\frac{M_{\mathrm{CaCO_3}}}{M_{\mathrm{Mg}}} = \frac{100.1\ [\mathrm{g/mol}]}{24.3\ [\mathrm{g/mol}]} = 4.1$$

$$CaCO_3[\mathrm{mg/l\ or\ PPM}] = 2.5 * C(Ca)\ [\mathrm{mg/l}] + 4.1 * C(Mg)[\mathrm{mg/l}]$$

In Finland, the hardness is expressed usually in German degree of hardness °dH (deutche Härte). Other units and ways to express the hardness of water are also used, for example French degree or ppm. The Table 1 shows the conversion and relations between different units.

Table 1. The unit conversion table from one hardness unit to another [4, p.216]

	mval/l	German degree °dH	French degree °fH	English degree °eH	ppm ($CaCO_3$)	mmol/l
mval/l	1	2.8	5	3.5	50	0.5
German degree °dH	0.36	1	1.78	1.24	17.8	0.18
French degree °fH	0.2	0.56	1	0.7	10	0.1
English degree °eH	0.29	0.8	1.44	1	14.3	0.143
ppm ($CaCO_3$)	0.02	0.06	0.1	0.07	1	0.01
mmol/l	2	5.6	10	7	100	1

When the numeric value has been calculated for the hardness, the water can be classified, for example, soft or hard. Classification includes several stepwise categories. There are some differences between the categories depending on the literature source. According to Vesikirja [5], the hardness is classified by the following way that is presented in Table 2. The hardness is divided in five different groups and the corresponding hardness values in °dH.

Table 2. Classification for water hardness [5, p.29]

Classification	**Scale**
Very soft	0 – 2.1 °dH
Soft	2.1 – 4.9 °dH
Moderately hard	4.9 – 9.8 °dH
Hard	9.8 - 21 °dH
Very hard	over 21 °dH

The geological environment is reflected in the water quality. In Finland the soil is mostly acidic and this leads to the water to be soft. The hardness varies in Finland according to the geological formations, and locally there might be some changes in the hardness within the seasons. Only in few places in Finland water is moderately hard or hard. [6]

For water consumers water hardness affects, for example, the usage of soap. The higher the water hardness is the more soap is needed to get a good washing result. Nowadays the washing products are not as easily affected by the hardness as before. Many of the wash powders have dosing guides in the baggage label for the different hardness of water. Most water companies provide quality information of the water they are supplying at their websites, where the consumers can easily check their water hardness and other relevant information that may affect their everyday life. Meri-Lapin Vesi also provides this information to their clients on their web page.

The high calcium amount of hard water will accumulate in the household appliances. Accumulation needs to be taken into account when using and cleaning machines. Also, if the hardness is too high, water starts to taste unpleasant. The above mentioned effects are the main reasons why hardness needs to be removed from water. Meri-Lapin Vesi has tips on their web page how to maintain the good condition of the household appliances with hard water, making them to last longer.

Having too soft water is problematic too. Moderately hard water will form a slight protective layer in the pipes. Soft water in the other hand is corrosive. Soft water enhances the pipes' and pipe instruments' metals to dissolve in the water. This lowers the water quality; as a result the quality requirements at the consumer's end might not be met anymore. [7]

World Health Organization states that hardness is not a health concern in the amounts found naturally in waters. People have different tolerance levels of tasting hardness, and the higher the hardness gets, the more unpleasant it is usually considered to be. The hardness of the may affect the taste and whether or not it is accepted as drinking water. [8] Finland's ministry of social affairs and health does not define hardness levels for drinking water in their quality parameters. The ministry states that water quality should be such that it does not cause problems to human health. Also water should not be corrosive or precipitate in the pipes. [9]

Meri-Lapin Vesi has high hardness in some of their groundwater reservoirs. The hardness may be up to 13 °dH in these reservoirs. The water reservoir used in this test run had a hardness of about 9-10 °dH. Meri-Lapin Vesi would like the water hardness to decrease to a level of 1 °dH for the permeate of nanofiltration. The permeate would then be mixed with water that has not been nanofiltered. The total hardness of the product water would lower when the waters are blended. The aimed hardness for the product water was 6-7 °dH.

2.2 Nanofiltration

Nanofiltration is part of membrane filtration technology. There are four different pressure driven membrane types, which are reverse osmosis, nanofiltration, ultrafiltration and microfiltration. [10, p.15.2] Membranes are semipermeable, which means that some particles cannot get through the membrane but that water and some particles will flow through the membrane. The difference between the four membranes types are the size of the particles they are able to filtrate. Also some of the membranes might have positive or negative surface charge and are able to filtrate then the opposite charged ions. The following Figure 1 shows the filtration abilities of the four different membranes. Due to membrane's pore size and charge, nanofiltration removes divalent ions. This makes nanofiltration an ideal process for removing hardness because hardness is caused by divalent ions.

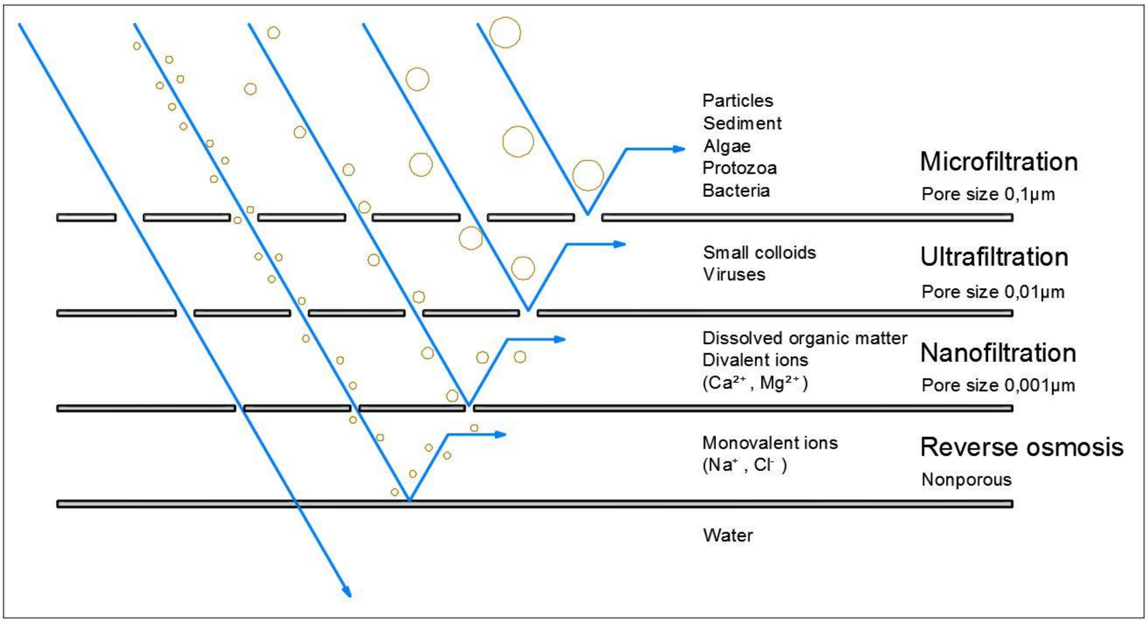


Figure 1. The different membranes and their pore sizes. List describes what each membrane is able to remove. [11 p. 957]

By the end of the 20th century, nanofiltration had made its breakthrough in the drinking water industry. Nanofiltration membranes were originally designed to remove hardness from water, but since the 20th century, their use for removing other components, such as nitrate or pesticides, has also been studied [12]. The removal properties for nitrate, pesticides and several other compounds have shown to be promising and also better than expected. Nanomembrane has proven to be good at removing natural organic matter, but not in as high amounts as the reverse osmosis membrane does. [13]

Nanofiltration membranes for drinking water treatment are normally made of synthetic organic polymers. Membranes can be cellulosic acetates or noncellulosic. Noncellulosic can be polyamides, polyurea, sulfonated polyfurans or other composites. [10, p.15.5] The membrane structure is usually asymmetric, no matter if it is made from one substance or composite of several materials. The surface layer of the membrane is normally denser than the layers beneath the surface layer. The separation process happens on the surface layer. After the surface, there is a more porous layer that allows the water flow better than the surface layer, and after that, there is a support layer that holds the membrane together. [14, p 25]

The membrane allows water and possibly small particles to go through its pores. The membranes have specific pore size and they are also presented earlier in Figure 1. Particles that are smaller than the pore size will most likely flow through the membrane with the water. Bigger particles than the membrane pores will be captured by the membrane.

The membranes can be laid in the filtration unit in several forms, for example, hollow fibres, flat sheets or spiral wounds. The hollow fibre is a tube having a small outside diameter and is said to be the most common configuration in micro- and ultrafiltration membranes. Flat sheets are single-layer membranes and more common in small laboratory scale. The spiral wound has several flat sheets stacked in layers and rolled around the collection tube. The spiral wound is the most often used configuration in nanofiltration. [11]

A spiral wound membrane configuration is presented in the Figure 2. Membranes are in an envelope setup and rolled around the central collection tube. Different layers of membrane and the envelope setup help to keep the filtrated and rejected water separate from each other. The membrane envelope has a spacer inside to allow the feed water to flow in between the membranes. The water filters through the membrane envelope and flows to the centre in to the collection tube. The concentrated feed water is prevented from entering the collection tube by the closed end of the envelope and is guided to exit the membrane unit separately from the permeate. This spiral wound is placed inside a closed vessel forming the filtration module.

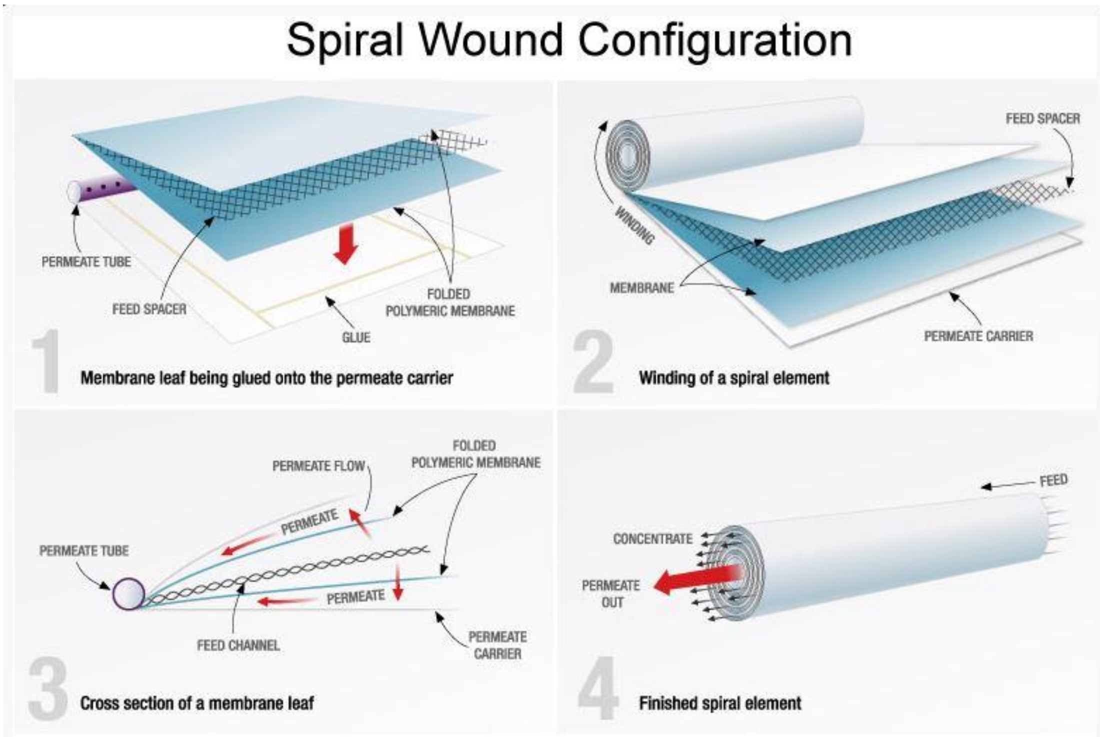


Figure 2. Configuration of the spiral wound membrane unit. [15]

When the feed water flows parallel to the membrane, it is called cross flow. Feed water is entering the membrane module with pressure, and with the help of this pressure the water will flow through the membrane. Filtered water will flow towards the middle of the spiral to the collection tube. Not all of the water that is fed in to the system will flow through the membrane. This water will continue to flow parallel to the membrane and will exit the unit as a concentrate.

Quite often nanofiltration is added as a part of already existing treatment plant to improve the process. Membrane filtration is most often added towards the end of the treatment process. The raw water quality determines the pretreatment process, i.e. what is needed to be done before the water can be fed to the nanofiltration unit. If the quality of the groundwater is good enough it might be enough to have nanofiltration as the only treatment for the water. For surface water, more extended pretreatment is needed.

2.2.1 Permeate

The end product that has filtered through the membrane is called permeate. Permeate is the wanted product water. The quality of the permeate depends on the membrane that is used. Even between the nanofiltration membranes there are differences in how big or small particles they are able to remove.

Molecular weight cut off (MWCO) is a number that describes the membranes ability to remove particles. This number describes the size of the particles of which most will be stopped by the membrane. Unit for molecular weight cut off is Dalton (Da). [14]

Most often the permeate quality is too good; therefore, it cannot be used as it is. This is because it might be demineralized or the hardness gets lower than wanted. Due to the low hardness of the permeate, only part of the water produced in the plant needs to be filtered with nanomembranes. The permeate can then be mixed with the water that is not nanofiltered to get the desired water quality. [6]

2.2.2 Reject/concentrate

The particles and the water that does not filtrate through the membrane are called reject. Basically the reject is concentrated feed water, and that is why it can also be called concentrate. The cross flow of the feed water will help to keep the membranes cleaner. Feed water cross flow will prevent, to some extent, impurities from accumulating on the membrane surface [11]. The parallel flow will “wash” the membrane surface and take the particles out from the membrane unit with the reject.

Rejection is a percentage for expressing how much a specific substance is rejected from the feed water. The final concentration of the substance in permeate is compared to the initial concentrate in the feed water. This ratio can be used to determine the percentage of how much from this substance is removed. The rejection can be expressed with the following formula [11]:

$$R = 1 - \frac{C_\mathrm{P}}{C_\mathrm{F}} * 100\ \%$$

Where: R = Rejection (%)

C_P = Concentration of the permeate [mg/l]

C_F = Concentration of the feed water [mg/l]

The concentrate has chemical composition as the feed water but in higher concentration. The concentrate disposal method depends on the concentrate content. If the concentrate is good enough, it can be ejected back to the nature. This varies highly according to the local laws, regulations and authorities. If the concentrate is not allowed to be ejected back to nature or there is not a possibility for it, it has to be lead to sewers to the waste water treatment plant. The concentrate can also be recirculated back to an earlier stage in the water treatment process in the plant. Sometimes the reject disposal method can be expensive if some special arrangements are needed. These kind of methods are, for example, deep injection well or evaporation.

2.2.3 Fouling

During filtration the membranes start fouling which means that the membrane's performance is getting poorer. Fouling means that the membrane surface either starts collecting film (scale) or the pores of the membrane will clog. Fouling leads to weaker permeate production. It is important to prevent fouling to reach optimum usage of the membranes and keep them in good condition for longer time. Fouling might be caused by a few different reasons. The following list presents some of them [14].

- Scaling of inorganic matter
- Colloidal fouling
- Organic fouling (absorption of organic molecules)
- Biofouling (microbial growth)

The bigger particles in the feed water might get stuck on the pores that have smaller diameter than the particle itself. Sometimes the parallel feed water is able to remove the particles from the pores. Particle that is stuck in the pore collects other particles around it. This is how a cake will form over the membrane surface. Also smaller particles can adsorb to the membrane surface or in the pores forming a coating layer. [11, p.984] Figure 3 shows these different ways of the membrane fouling. Sometimes the gravity plays its role in fouling by allowing the particles to settle on the surface, but this is often avoided by installing the membranes vertically.

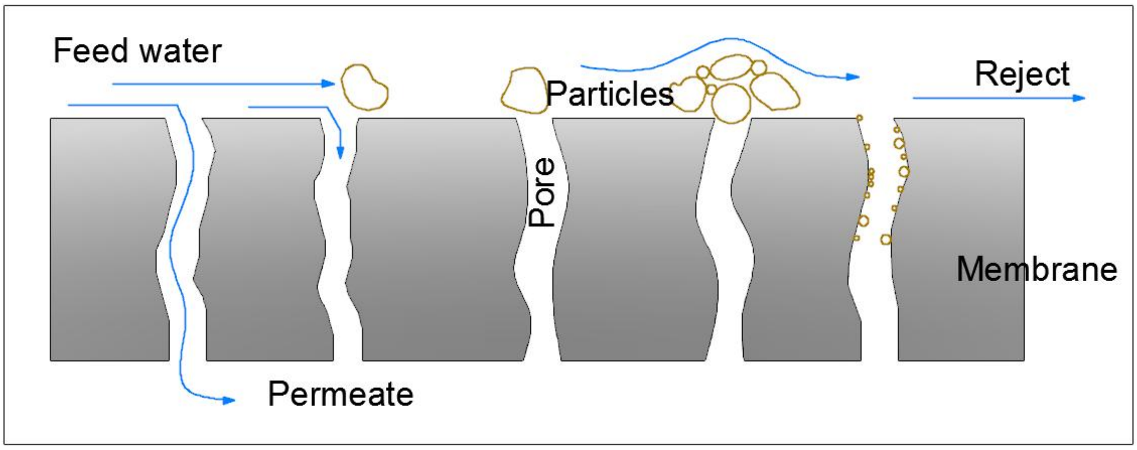


Figure 3.Fouling of the membrane pores. Water will flow through the membrane pores. Particles get stuck on the pores or adsorbs on the surface.

It is often necessary to have some pretreatment before the membranes to remove the bigger particles from the water. Pretreatment could be cartridge or bag filtration. Also to prevent fouling an antiscalant can be added to the feed water.

The pressure drop on the permeate tells to the plant operator that the membrane is clogged. This indicates that it is time to wash the membranes. Membranes may be washed with chemicals or commercial cleaning solutions. Purpose of washing is to open the pore clogs and remove the film from the membrane surface. In the course of time, the fouling of the membrane becomes irreversible and the washing does not help. In this case, the membranes have to be replaced with new ones.

2.2.4 Antiscalant

Antiscalant is a chemical used for preventing the fouling of the membrane. Different antiscalant chemicals will prevent the scaling of different substances. Scaling appears when the concentration of the substance is so high that not all of it is in soluble form anymore. [14] Particles will form bigger colloids or crystals together. Then the pores are in the risk of being clogged by the colloids. Antiscalant will prevent this formation of colloids or crystals. When the ions cannot scale, they will be washed away with the concentrate, instead of clogging the pores of the membrane.

Antiscalant is usually added into the process before the prefiltration so that the chemical has enough time to mix in the water. The prefiltration adds a physical barrier to the flow causing small momentarily turbulence in the flow. This helps the antiscalant to mix evenly in the water before the water reaches to the membranes.

2.2.5 Recovery and flux

Recovery and flux are often mentioned when membrane performance is discussed. They are values used for describing the effectiveness of the membranes. Recovery describes the ratio of how much is the permeate production from the total water that is fed into the membrane unit. Recovery is presented in percentages and can be calculated when the volume flows of the process are known. The common formula to calculate this is the following: [10]

$$Recovery\ [\%] = \frac{Q_{\mathrm{p}}\ [\mathrm{volume/time}]}{Q_{\mathrm{f}}\ [\mathrm{volume/time}]} * 100\ \%$$

Where: Q_p = the flow of the permeate

Q_f = the flow in to the filtration system

The recovery tells how much of the feed water is turned into the end product permeate. This value is important for monitoring the membrane's effectivity and overall nanofiltration performance. Recovery helps to monitor any possible changes that might happen in the process. If the recovery rate of the membrane drops, it might be a sign of the fouling of the membrane or the feed pressure not being high enough to create enough permeate.

Flux is a value that describes the membranes product water flow per active membrane area per time. Flux is often expressed as l / m^2h. This value is also used to describe the membrane performance. The value describes how many litres one m^2 of membrane is able to filtrate in an hour. The formula to calculate the flux of the nanofiltration is as follows: [16, p 42]

$$Flux = \frac{1000 * Q_P}{A}$$

Where: Flux = l/m^2 h

Q_p = flow of the product (m^3/h) (1000 is correction for l/h)

A = area of the nanomembranes (m^2)

Recovery and flux are used to describe the membrane's abilities in the given condition. Thus, if the recovery is presented to be 80%, it means that from 1 m^3 water that is fed in the membrane unit, 0.8 m^3 is transferred to permeate and 0.2 m^3 is reject. If the pressure or the feed water quality changes, the recovery and flux can change. Then the flux and recovery can be different from those at the start of the filtration process.

2.2.6 Critical pressure and flux

Critical pressure and flux describes the situation when particles start scaling on the membrane surface. In the optimum condition the cross flow prevents the particles from scaling on the membrane surface. Scaling cannot be fully prevented, but with the correct dimensioning of the system, the scale formation is small on the membrane surface. Particles accumulate on the membrane surface more if the filtration is operated with too high pressure. Usually increasing the pressure aims to increase the flux, but this has opposite effects than desired. When the particle cake formation has built up on the surface, it lowers the flux. [17]

The flux normally changes after the filtration process has started. The flow and the feed water particles find their balance on the surface of the membrane. This is seen as small decrease in the flux. If the feed pressure is increased in this situation, the balance is disturbed and cake formation grows until new balance has been found. Operating over the critical pressure or flux, the membrane fouling is most likely to happen faster than under critical values. This will decrease the life time of the membrane. [17] Typical operating flux for the membranes varies between 22-27 l/m²h. Also the typical operating pressures are 3.4-10.3 bar. [10]

2.3 Cost of nanofiltration

Cost of nanofiltration is affected by several different variables. The investment and operational costs determine if nanofiltration is a feasible option compared to the traditional hardness removal methods. These also define the possible increase for clients in the water price. The investment costs are getting lower throughout the years. Still, in the 1990’s nanofiltration was considered as a huge investment compared to the traditional treatment methods [18].

2.3.1 Cost parameters

The total cost of nanofiltration will be determined by different parameters. Each nanofiltration case is individual. Nurminen [19] studied the operational cost of nanofiltration in three different water treatment plants in Finland. These nanofiltration plants did not necessary use nanofiltration for removing hardness. The parameters affecting the investment costs were listed as follows: [19, p.29]

- raw water procurement
- raw water pretreatment
- membrane filtration process, chemical feed and monitoring
- permeate treatment
- permeate feed to the network
- concentrate treatment
- concentrate byproducts treatment
- concentrate removal
- process and monitoring facilities
- site and soil modifications

When nanofiltration is added to the existing process to improve the process, some of the parameters listed above already exists and will not affect the cost calculations. In Meri-Lapin Vesi’s case, the plant already exists, and the water procurement and pretreatment are factors that do not need be taken into account. Concentrate treatment is also a case sensitive factor. On the basis of the previous tests in Meri-Lapin Vesi, the concentrate quality was good enough to be released in to the nearby river. Some nanofiltration treatment plants in Florida utilize deep injection wells to dispose of the concentrate. [20]This kind of systems increases the investment on the nanofiltration.

The investment cost of nanofiltration has come down during the years. The manufacturing costs of membranes are cheaper now than in the early years of membrane technology. This decreases the operational cost of the filtration since the replacing of the membranes is not as expensive as it used to be.

Some of the operational cost factors affect the total cost more than the others. Nurminen [19] also examined two different studies to determine how the operational costs were divided among different parameters. Table 3 presents the results of these two studies. The values are presented as percentages from total operational cost. Six main cost factors presented in these studies were energy, replacing membranes, work, chemicals, prefilters and spare parts. Also Van der Bruggen et al. [12] has presented how the operational costs were divided in their study. This cost division is presented in the pie chart in Figure 4.

Table 3. Summary of operational costs from Coté's and Bergman's studies. [19]

Parameter	**% of the operational costs**	
	Coté	**Bergman**
Energy	25	29
Replacing the membrane	26	10
Work	29	31
Chemicals	13	17
Pre filters (cartridge)	4	13
Spare parts	3	
Total	100	100

Energy and the work force are the factors that contribute the most for the total operational cost. Chemicals (antiscalant and cleaning chemicals) are also major cost factors. These six parameters mentioned in the table are the main parameters studied in almost all the studies. Some other regional factors are found to contribute on the total price of the nanofiltered water. For example taxes related to water usage or waste water generation. [12]

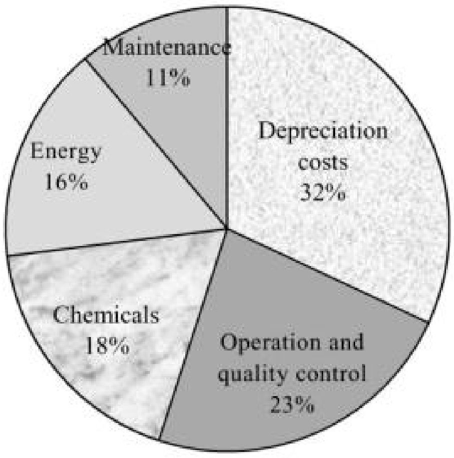


Figure 4. Summary of the operational cost based on Van der Bruggen's study. [12]

Nanofiltration is case sensitive. In some studies the membranes did not scale easily, and they did not need to be washed often. In such cases, the chemical costs were smaller than in cases where the membrane needed to be washed frequently. In one study, the antiscalant was found to cause fouling of the membrane rather than preventing it [16]. It is also important to find the right kind of membrane for the need. Membranes have different pore sizes and are designed for different purposes. If the membrane is not suitable for the need, the cost will start to accumulate. The membrane might scale more, need more cleaning or chemicals, and then the use of nanofiltration is not cost efficient.

2.3.2 Formulas used in cost calculations

The operational costs of nanofiltration are usually calculated for the full size plant per annum. The total cost is composed of small individual parts that have been discussed above in part 2.3.1. From the total annual operational cost, the price for 1 m³ of permeate can be determined. The formulas presented here were used for calculating the operational costs for Meri-Lapin Vesi Oy.

The membranes usually have approximately a 5-year life expectancy. Not all of the membranes are changed at once. Normally 20 % of the membranes are changed during one year. Cost for replacing the membranes can be calculated with the following formula:

$$C_{\mathrm{M}} = M * 0.2 * €_{\mathrm{M}}$$

Where: C_{M} = cost of membrane replacement €/a

M = amount of membranes in the plant

0.2 = 20 % membranes replaced annually

$€_M$ = price of one membrane

The cost of the cleaning is a sum of the chemicals used for washing and energy used for heating up the washing water. Also discharging waters into the sewers adds up to the cleaning costs. The following formula can be used for cleaning calculation:

$$C_{\mathrm{W}} = (C_{\mathrm{WW}} + C_{\mathrm{H}} + C_{\mathrm{C}}) * X$$

Where: C_W = cost of washing €/a

C_{WW} = cost of waste water discharge

C_H = cost of heating up the washing water

C_C = cost of chemicals

X = number of washings done in a year

$$C_{\mathrm{WW}} = V * €_{\mathrm{WW}}$$

Where: C_{WW} = cost of waste water discharge €/wash

V = volume of the discharged waste water in m³

$€_{WW}$ = price of wastewater €/m³

$$C_{\mathrm{H}} = E_{\mathrm{H}} * €_{\mathrm{E}}$$

Where: C_H = cost of heating of the wash water

E_H = electricity used for heating the wash water

$€_E$ = price of electricity €/kWh

The price of antiscalant can be calculated for the whole year when the consumption of the antiscalant is known. The antiscalant is fed to the feed water. The antiscalant dosage is scaled by the feed water.

$$C_{\mathrm{A}} = D * Q_{\mathrm{FA}} * €_{\mathrm{A}}$$

Where: C_A = cost of antiscalant €/a

D = antiscalant dosage l/m³

Q_{FA} = feed water in a year m³

$€_A$ = price of antiscalant €/l

Energy is consumed by the pumps of the feed water and antiscalant. The following formula can be used to calculate the cost of energy of nanofiltration system.

$$C_{\mathrm{E}} = E_{\mathrm{P}} * Q_{\mathrm{FA}} * €_{\mathrm{E}}$$

Where: C_E = cost of using energy €/a
E_P = electricity used for pumping feed water kW/m³
Q_{FA} = feed water in a year m³
$€_E$ = price of electricity €/kWh

2.4 Nanofiltration in the world and in Finland

Nanofiltration is not widely used in Finland and especially for removing hardness. There are several treatment plants that have nanofiltration as part of their treatment process, but they are not using it for removing hardness as Meri-Lapin Vesi would be using it. Hardness removal studies have been conducted in many other countries. Throughout the years, several pilot scale studies have been done for testing the removal properties of the nanofiltration system. These studies have been conducted all around the world and some of the results and user experiences will be discussed in the following paragraphs.

Several studies concluded that the operational costs are smaller when the capacity of the treatment plant is higher. Some of the studies were done in 1990's and the costs are converted into today's value. This does not always compare and tell the full truth. The nanofiltration technology itself has become cheaper and this might reflect in the operational costs for example regarding the replacing of the membranes.

2.4.1 Germany

A pilot test conducted in Germany in 2002 was done to study the removal of hardness and natural organic matter (NOM) from groundwater. Operational cost calculations were made to see how much nanofiltration would affect the price of treated water. The test showed good results for removing NOM and hardness. According to the results, the hardness removal rate was better than what the manufacturer had announced for the membrane. [21]

The study was made in a pilot scale with a spiral wound nanomembrane. In the test, there was a reject recirculation option. Pressure was adjusted with a bypass and concentrate valve. The membrane used in this study was NF200B by FilmTech. The test was made for three different reject recoveries. The cross flow and pressure was kept the same in all three recoveries. During the test the membrane did not suffer from significant fouling and this gave high lifetime expectancy for the membranes.

Table 4. Parameters used for calculating the operational costs of nanofiltration in Mainz, Germany. [21]

Mainz, Germany Parameter	Unit	
Capacity	m^3/a	7 300 000
Membrane		NF200B, FilmTech
Recovery	%	85
Operating pressure	bar	5.5
Raw water hardness	°dH	18.5
Permeate hardness	°dH	7.92

Final cost calculations were made for two different capacities. Higher capacity was calculated so that all of the water in the plant is treated with nanofiltration. Parameters used in this calculation are presented in Table 4. In lower capacity (2 850 000 m^3/a) only part of the water in the plant is nanofiltered and then mixed with traditionally treated water.

According to this study, the operational costs of membrane filtration is a combination of pre-treatment, capital costs, energy, membrane replacement, maintenance, chemicals, concentrate disposal and post treatment. The increase in the price of water for consumers was estimated to be 0.23 €/m^3 permeate produced for the full nanofiltration capacity. For the smaller capacity, the price would increase to 0.27 €/m^3 permeate produced. Nevertheless, the price increase for consumers would be only 0.11 €/m^3 since the permeate would be blended with the traditionally treated water. [21] The prices presented are in euros from year 2002 and do not take inflation into consideration.

2.4.2 Belgium

In 2001 Van der Bruggen et al. [12] studied in a laboratory scale nanofiltration for removing pesticides, nitrate and hardness. This study also focused on cost evaluation of the filtration. Tests were done with several different nanomembranes. The hardness removal properties of the membranes gave good results and also the nitrate removal rate was better than expected. The pesticides removal was dependent on the membrane, and some membranes performed better than others. After nanofiltration, the water should be blended with non-nanofiltered water so that the hardness level would not drop under the desired level.

The tests were carried out in laboratory scale with groundwater provided by Flemish water company WMW (Vlamise Maatschappij voor Watervoorziening) in Belgium. Four different nanomembranes from two different manufactures were tested. Membranes were selected from DOW/FilmTech (NF70 and NF45) and Toray Ind. Inc. (UTC-20 and UTC-60). The final economical evaluation was conducted with the NF70 membrane. This membrane was chosen on the basis of its removal results. The cost calculations were done with several different pressures to find the optimum conditions for the nanofiltration. The parameters for calculating the operational costs are presented in Table 5. In this table, the operational pressure is 8 bar which is the optimum pressure for the filtration system giving the minimum operational costs. Calculations were also made with 5, 10, 15 and 20 bar.

Table 5. Parameters used for calculating the operational cost of the Belgium nanofiltration plant. [12]

Parameter	**Unit**	**Belgium**
Capacity	m³/a	17 500 000
Membrane		NF70 8040, DOW/FilmTech
Recovery	%	80
Operating pressure	bar	8
Raw water hardness	°dH	15.7
Permeate hardness	°dH	0.787

Operational costs were calculated with energy and chemical consumption, maintenance and specific operation costs. Additional costs for nanofiltration come from taxes that have to be paid for discharging the concentrate in the sewers and using groundwater. Operational costs calculated for a capacity of 2000 m^3/h was 0.17 €/m^3 permeate produced, including taxes. With a 10 times lower capacity, the cost would increase up to 0.26 €/m^3 permeate produced. [12] All the prices presented are in euros from year 2001 and does not take the inflation into consideration.

2.4.3 USA

In Florida, the USA, several groundwater treatment plants were using nanofiltration already in the 1990's. Bergman [20] studied in the mid 1990's the construction and operational costs of treatment plants with nanofiltration. Generally, the results showed that the operational costs are smaller when the capacity of the plant is higher. All in all, this study mentioned 14 operational nanofiltration plants and 5 plants that are under construction. Operation and maintenance costs were collected from 7 of the operational plants. The operational costs were presented as just the costs of nanofiltration without final blending of the water. All of the plants used hard groundwater. [20]

The operational and maintenance cost from Bergman's study of nanofiltration plants in Florida mid 90's are presented in Table 6. The operation capacity of the plant is also presented in the table as m³ permeate production per day. The costs are presented as euros per m³ permeate produced.

Table 6. Operation and maintenance cost of nanofiltration plants in Florida. [20]

	Capacity m³/d	Chemical €/m³	Energy €/m³	Work €/m³	Membrane replacement €/m³	Other €/m³	Total €/m³
Plantation-Central plant	45 400	0.04	0.04	0.04	0.026	0.013	**0.158**
Fort Mayers	45 400	0.026	0.053	0.066	0.026	0.04	**0.198**
Collier County	45 400	0.026	0.053	0.053	0.026	0.026	**0.198**
Indian River County South	22 700	0.04	0.079	0.053	0.013	0.013	**0.198**
Dunedin	22 700	0.066	0.066	0.053	0.04	0.026	**0.251**
Boynton Beach	15 100	0.013	0.04	0.132	0.026	0.013	**0.238**
St. Lucie West Development	3 800	0.119	0.224	0.198	0.04	0.132	**0.7**

The costs in the Table 6 have been converted from 1996 US dollars to 2016 euros. Factor 1.51 was used to convert the 1996 dollars to 2016 dollars. [22] The factor 0.88 was used for converting US dollars to euros. This factor was European Central Bank's rate for the US dollar in April 2016. [23]

On the basis of the information given in Table 6, it can be concluded that the high capacity plants have much lover operational costs. In a low capacity plant treating 3 800 m³/d, the operational costs was 0.7 €/m³ permeate produced. This is almost three times higher cost than the costs for the next highest plant with a much higher capacity. For higher capacity plants treating 15 100 – 45 400 m³/d, the operational costs were 0.16 – 0.25 €/m³ permeate produced.

2.4.4 Finland

Nurminen [19] studied three different water treatment plants in Finland and estimated the cost that the adding of nanofiltration to the process would cause for these plants. All three studied plants were using groundwater, but the nanofiltration plants were not designed for water softening. These three plants have a smaller capacity than any of the previously mentioned studies conducted in abroad (Germany, Belgium or USA). Nanofiltration capacities varied from 150-700 m³/d. In all of the plants, nanofiltered water was mixed with traditionally filtrated water.

The prices for permeate produced varied between 0.18-0.26 €/m³. When nanofiltered water was blended with non-nanofiltered water, the total expenses for the water were lower. The costs for blended water varied between 0.03-0.15 €/m³. The operational costs that Nurminen got for the plants have been converted from Finnish mark to euros. The factor for the conversion is from Statistics Finland's factor for the value of money [24].

The three different studied plants in Finland were in Kempele, Mustasaari and Laitila. In all of the plants different FilmTech nanomembranes were used. Each case was unique and different. In Kempele, the membranes were washed with acid once in a month. In Mustasaari the membranes were washed daily with base solution and once a week with acid. In Laitila, during 1-year-4-month operation, the membranes were washed only twice with acid and base. [19] These kinds of differences reflect quickly in the chemical consumption and operational cost.

Also in Espoo City Waterworks, a nanofiltration study has been conducted. The nanofiltration was tested for improving the existing surface water treatment process by removing natural organic matter. Pilot scale studies were conducted in December 1999 to February 2000. Four different operational pressures were chosen and two different recoveries to study. A capacity of 18 000 m³/d was used for a full-scale plant estimation. The operational costs were calculated to be 0.103-0.112 €/m³ permeate produced. [25]

2.4.5 Others

In 1998, a study titled *Performance of 3 years' operation of nanofiltration plants* was done by Gaid et al. [18] Three different nanofiltration plants were studied in France and Great Britain. This study again showed that nanofiltration has a good ability to remove hardness. However, membrane processes were considered more expensive than traditional purification methods. Energy consumption and membrane life span were the main issues for increasing the price of membrane filtration systems. [18] For Mery-sur-Oise plant in Paris the operational costs was 0.12€/m³. [12]

In 2009, a study in Morocco was conducted to economically evaluate fluoride removal from groundwater with nanofiltration. The calculations were made based on the previous studies made on the same topic. The design criteria were a capacity of 2 400 m³/d with an 84 % recovery rate and a 97.8 % fluoride rejection rate. Total operating costs were calculated to be 0.212 €/m³ permeate produced. This study concluded that the capital costs were calculated to be much higher when using a model than using the real data. The operational costs were still comparable even though the operating costs based on real data were bit higher than model-based costs. Based on the model, the operating cost was 0.16 €/m³ permeate produced. [26]

Removing organic matter from river water was studied in Valada, Portugal in 2006. The experiments were made in lab scale, and the results were used to create a model to calculate the costs for the operation of the plant. The capacity of the modelled plant was 100 000 m³/d and the modelled cost for the treated water was 0.214 €/m³. [27]

2.4.6 Summary of cost of nanofiltration around the world

The studies are not fully comparable with each other. This is due to the fact that they have been conducted in different countries and within a large time span. The cost comparison is indicative of how different parameters might affect the costs. Newer updated cost information and studies are not as easily accessible as older studies.

During the years membrane technology has evolved and, for example, producing membranes has become cheaper. This is seen as a price decrease in the membrane replacement. Also the energy prices are not comparable because the price of energy varies in each country. Also national or local regulations might cause extra costs.

Few general conclusions can be drawn from the studies that have been performed around the world. The plant capacity affects the cost of the permeate. When the permeate production is higher, the cost of producing it gets lower. The quality of the feed water affects the price. When more antiscalant or cleaning is required to maintain the membrane performance, the price increases.

The prices of nanofiltered water and plant capacities are collected into a chart that is presented in the Figure 5. Each dot in the chart represents one individual plant or a result of a study. The plants in the USA are presented as red diamonds. The US plants are groundwater plants, and nanofiltration is for removing hardness. The orange diamonds represent studies where nanofiltration has been used for removing hardness from groundwater. Green dots represent the surface water plants. Purple dots are groundwater plants where nanofiltration has not been used for hardness removal. In general, the diamonds are groundwater plants with hardness removal, and dots are plants using nanofiltration for other than harness removal.

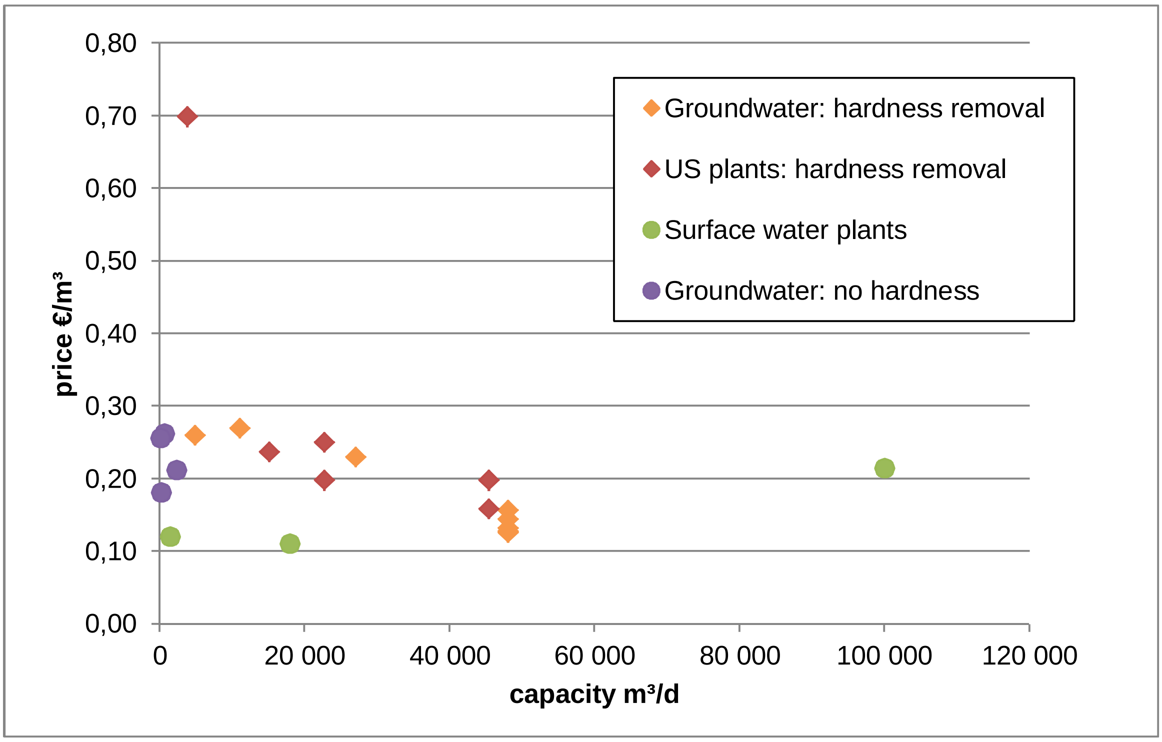


Figure 5. Price of permeate vs. plant capacity. Price and plant capacity compared from the different studies. [12, 19, 20, 21, 25, 26, 27]

The price range seems to be small for most of the plants. The one high price (0.7 €/m³) is the small capacity plant presented in Bergman's study. It was the only small-capacity plant whose price was presented in the study. The study was conducted 20 years ago, and the prices are not directly comparable to more new ones. Today the price in this plant might be lower and closer to those of the other plants presented in the chart. The price decrease with the plant capacity was much more drastic in Bergman's study than in the other studies.

The plants discussed in this section are also presented in Appendix 1. The design parameters and the costs for different parameters are compiled to a table. The prices have been transformed into euros as described earlier in section 2.4.

3 Practical part

In Meri-Lapin Vesi's previous study, four different test runs were executed with two different membranes. In these test runs the fourth test was considered as the most successful. The membrane that was used in the fourth run had the highest removal rate of hardness. It was even higher removal than what manufacturer informed as the removal rate for the membrane. However, the test was not fully a success because the permeate production was lower than expected. This was assumed to be due to the too low feed water pressure. The pump used in previous study did not provide enough pressure for the feed water. Also iron and manganese were suspected to cause the fouling of the membrane lowering the permeate production.

It was decided that new test runs would be made with the most promising membrane of the previous study. Aim of the new test runs was to test two differently pretreated waters with the membrane to see how it suits for these two different situations. In new test runs the aim was to eliminate the factors that previously caused the low permeate production. The idea was to wash the membranes 1-2 times during the test runs. This would show how the washing affects the membranes and what the chemical consumption is during washing. The operational cost is defined based on these new test runs.

BWT Separtec Oy has played an important role in the test runs. They provided the nanofiltration membranes and the test unit to Meri-Lapin Vesi. BWT Separtec Oy is an international company providing water purification solutions for household, municipal and industrial scales. BWT Separtec took part in the previous test runs that were conducted in Meri-Lapin Vesi.

3.1 Test run configuration plan

The current treatment process for groundwater is iron and manganese removal process where dynasand filtration is the last stage of the process. The plan for the tests runs was to test two different types of feed water. First test was planned with feed water that is filtered twice through the dynasand unit. Twice dynasand filtrated water is used so that iron and manganese is filtered as much as possible from the water. The second test was originally planned with only one time dynasand filtered water. This was chosen because if in the future the process would be changed to have a higher capacity. With higher capacity it is not possible to filter the water twice with dynasand. The aim was to see if the once dynasand filtered water would have good enough quality for nanofiltration. And would the membranes be fouling more than with twice dynasand filtered water.

The original plan was changed during the test. The first test was executed as it was planned, but the second test was altered. The water used for the second test was still twice dynasand filtered water, but an activated carbon filter was added to the process. Tests were still executed with two different water qualities.

The setup, how nanofiltration was placed in to the existing treatment plant, is presented in the process flow chart in Figure 6. The feed water for the nanofiltration was taken after the existing process. The current treatment is iron and manganese removal process and as a last stage of the removal process the water is sand filtrated in dynasand units. The feed water for nanofiltration was taken after the dynasand.

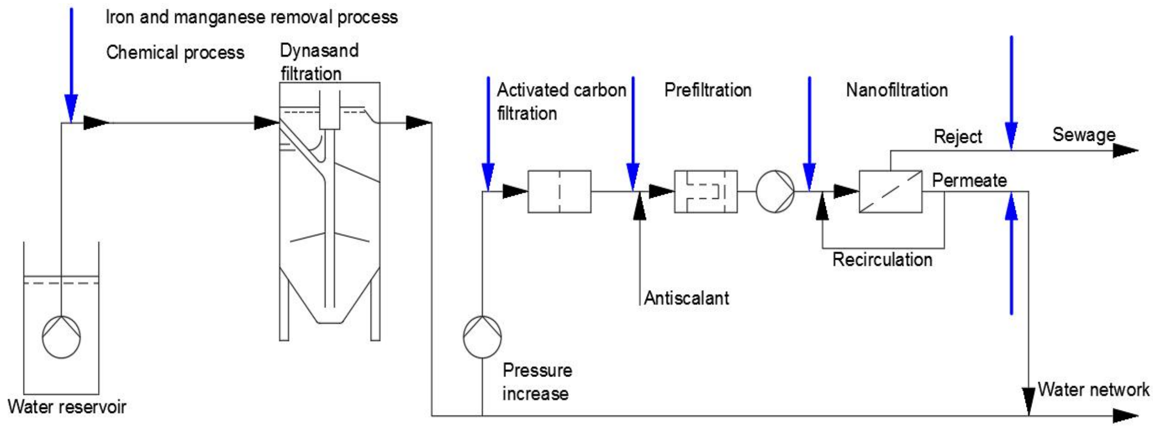


Figure 6. Process flow chart for the nanofiltration. How the nanofiltration unit is added in to the existing plant. Blue arrows indicate where from the process samples are taken.

For the nanofiltration process a pump was needed to increase the pressure because without the pump the pressure on the membranes is not high enough to produce enough permeate. Also too low feed pressure can stop the whole process. Higher permeate production was expected with more efficient pumping.

Antiscalant was added to the feed water after the pressure increase pump to prevent the fouling of the membranes. Antiscalant is added before prefiltration so that there is enough time for the chemical to mix properly in the feed water. Bag filters were used as a prefilter. Prefiltration was before nanofiltration to capture possible bigger particles that has passed through the existing process.

Activated carbon filtration was added to the process for the second test run. The purpose of the activated carbon was to remove natural organic matter that has passed through the already existing process. Natural organic matter causes fouling of the membranes and activated carbon filtration is preventing this to happen by removing the natural organic matter. Activated carbon filter was added in the beginning of the process after the pressure increase pump and before the bag filter.

Membranes were planned to be washed about 1-2 times during the test runs. Normally the pressure change and the drop in permeate production indicates the fouling of the membrane. In pilot test the cleaning might not represent the real full scale cleaning situation. This is because in pilot scale the membranes might not clog fully. This was not the case in this test run. During the test runs the membranes were flushed and washed twice.

A plan for taking samples for monitoring the water quality throughout the whole process was created. Blue arrows in the Figure 6 indicate the sampling points. Enough samples are required to follow any changes in the process or in the raw water quality. The sample test plan will be described more detailed in part 3.1.4.

3.1.1 Filtration unit and membrane

The most promising membrane used in previous test was chosen for the new test runs. This membrane was DOW FilmTech NF90-4040 nanofiltration membrane. The more detailed information of the membrane can be found from the membrane product sheet in the Appendix 2. Based on the manufacturer’s information this membrane is suitable for removing salts, iron, nitrate and organic compounds. The membranes rejection rate is said to be 97 %.

The manufacturer gives operating guidelines and limits for the membrane, such as the operating pressure or pH range. Table 7 presents the manufacturers requirements for the feed water quality. For the optimum operation of the membrane the feed water should fulfil these parameters.

Table 7. The feed water quality requirements for the membranes.

Parameter	Unit	Max value
pH		3-10
Temperature	°C	45
Permanganate value	mg/l	10
SDI		5
Fe^{2+}	mg/l	4
Fe^{3+}	mg/l	0.05
Mn	mg/l	0.05
Al	mg/l	0.05
Free Cl_2	mg/l	0.1

The filtration unit test rig was BWT PERMAQ® PRO 2550 Reverse Osmosis Plant. The test rig's product sheet is presented in the Appendix 3. The test rig has place for 12 membrane units. Two parallel lines have both 6 membranes in series. BWT Separtec has guided Meri-Lapin Vesi in the use of the membranes and test rig.

3.1.2 Antiscalant

The antiscalant that BWT Separtec recommended for the test runs was fumados SG. Manufacturer tells that the product is suitable for drinking water purpose and compatible with ANSI/NSF Standard 60 under 10 mg/l levels. Most of the chemical details of the product are trade secrets. For the advantages of this antiscalant the manufacturer lists its low feeding requirement, suitability for all membrane types and stability in all pH-values and temperatures.

3.1.3 Time schedule

The pilot size test is short small scale trial of the desired process. The goal of the pilot test is to demonstrate how the system would work. Based on the test runs, estimation for full a scale operation can be done. In such short test runs, samples needs to be taken often allowing the close monitoring of the process.

Table 8 shows the planned duration of the pilot test and the days when the samples were taken. Sample days were evenly spread in the work week for every other day.

The pilot rig arrived to Meri-Lapin Vesi in the beginning of the week, leaving few days' time to assemble the rig before the tests. The rig was assembled and up for running in Tuesday, even though the time schedule gave more time for initial starting.

Table 8. The duration of the pilot test and the sample taking days.

Week	Monday	Tuesday	Wednesday	Thursday	Friday
1			**S**		**S**
2	**S**		**S**		**S**
3	**S**		**S**		**S**
4	**S**		**S**		**S**
5	**S**		**S**		**S**
6	**S**		**No sample**		**No sample**
7	**No sample**		**No sample**		**No sample**
8	**No sample**		**S**		**S**
9	**S**		**S**		**S**
10	**S**		**S**		**S**
11	**S**		**S**		

15 times

10 times

The test was planned to be done with two different water qualities. First one with two times dynasand filtered water which is represented with blue in the Table 8. The green colour represents the second test, done with the activated carbon filter. The first test was planned to last 7 weeks with 19 sample days. But due to the changes made during the test, there were only 15 sample days. The second test was planned to be shorter, about 3 weeks with 10 sample days. The days marked with S in the table are the sample days.

A journal was also kept during the sample period. Each time changes were made or something significant happened, for example, test rig adjustments, error messages, samples and washings, that was marked down in the test journal. The pressures and flows of the system were recorded in the journal.

3.1.4 Sample plan

Water samples are necessary to take to follow and monitor the process. The sample plan was created to follow the possible changes and to see where these changes occurs. The sample points are represented as blue arrows in the Figure 6. Sample needs to be taken after each different stage of the process. This way each of the processes are monitored and the possible changes can be located from the process.

Table 9 represents the analysis carried out from each water sample. Samples were taken from raw water, feed water, activated carbon filtrated water, prefiltered water, reject, wash waters and permeate. Samples were taken according to the sample schedule in Table 8. Samples from wash waters were taken only when membranes were washed.

Table 9. Parameters that are analyzed from the water. Parameters marked with x and highlighted with grey were analyzed every time. Parameters with o were analyzed on Mondays.

Parameter	**Raw water**	**Feed water**	**Activated carbon filtered water**	**Prefiltered water**	**Reject/ wash water**	**Permeate**
pH	x	x	x		x	x
Conductivity	x	x	x		x	x
Alkalinity	x	x	x		x	x
Calcium	x	x	x		x	x
Manganese	x	x	x	x	x	x
Fe^{2+} (soluble iron)	x	x	x	x	x	x
Total iron	x	x	x	x	x	x
Magnesium	x	x	x		x	x
Permanganate value	x	x	x	x	x	x
Temperature	o	o	o		o	o
Carbon dioxide	o	o	o		o	o
Chloride	o	o	o		o	o
Silicate	o	o	o		o	o
Sulfate	o	o	o		o	o
Suspended solids	o	o	o		o	o
Turbidity	o	o	o		o	o
TDS	o	o			o	o
TOC	o	o	x	o	o	o
Total phosphorus					o	
Total nitrogen					o	
Odor		o				o
Taste		o				o
E. coli	o	o				o
Coliform bacteria	o	o				o

In Table 9, X represents analyses that were made each time the sample was taken. The extended analyses are presented with O. 9 different parameters were analysed from the samples every time the sample was taken. More extent analyses were made from Monday’s samples. From activated carbon filtrated water also total organic carbon was analysed every time.

3.2 The execution of the test run

The journal which the plant operator has held during the test is presented in the Appendix 4. In the following chapters the main events of the test runs are described. More detailed results of the test runs are discussed in part 4.

3.2.1 The first test run

The test runs were started at the beginning of the February on Tuesday 09.02.2016. After the first few days of operation, it was noticed that the permeate production was decreasing. With the guidance of the test unit supplier BWT, recirculation of the permeate was decided to add in the process. However, after the first weekend the recovery was still low. After one week, the first wash was decided to be done.

On Wednesday (10.02.2016) afternoon antiscalant feed was noticed not to work. The antiscalant feed was off, most likely over a day, in the beginning of the test. The fouling of the membrane in the beginning of the test was thought to be due to this malfunction of the antiscalant feed.

The first washing was done with acid and then twice with base solutions. The bag filter before the membranes was changed, since it was noted to be dirty. The washing of the membrane removed particles that the membranes had collected on them. This was visible to eyes, since the wash water was noted to be brown. The recirculation of the permeate was set in to maximum after the washing.

After the first wash, the permeate production was not on a satisfactory level. A new washing of the membranes was discussed. However, it was decided to keep the test running without washing. This was to see on what level the permeate production would settle.

A month after the start, a meet up was set with Meri-Lapin Vesi and BWT Separtec. In the meeting was discussed how the permeate production could be increased. At this moment the organic matter in the feed water is held as the main reason for the scaling of the membrane. The organic matter content in the feed water was at the maximum level of the feed water quality recommendation for the membranes. An activated carbon filter was decided to add in the filtration process to remove organic matter. It was also decided that the membranes will be washed with a strong commercial wash solution to remove as much dirt from the membranes as possible.

The sample taking was stopped at the end of the first test. This was decided because it was already know what the quality of the water was that was fed to the membranes. Likewise the quality of the permeate and reject were know, based on the samples taken and analysed. The journal was kept during the time, when samples were discontinued.

At the end of the first test the membranes were washed several times, to ensure they are as clean as possible. Washing was done with acid, base and commercial washing solution. The wash water was noted to be black during one of the washings. This was darker than any of the previous washings.

3.2.2 The second test run

New plan for the second test run was made. The feed water was twice dynasand filtered, but the activated carbon filter was added to the process. The aim was to see if the membranes will scale less, when the amount of organic matter decreases in the feed water.

Before the second test the membranes were washed. After the washing the activated carbon filter was added to the process before bag filter. The second test was started with as clean membranes as possible. Total organic carbon was analysed from activated carbon filtered water sample every taken sample, to monitor that the carbon filtration will work.

The second test was started on 30.03.2016. Soon after the start, some problems with the activated carbon filter was encountered. Some of the carbon got loose from the filter and pressure change indicated that the flow in the filter had changed. A counter current wash was executed to the filter.

A week later preliminary results of the sample analyses from the first week of the second test run came. The activated carbon filter was noticed to lower the amount of organic matter. Samples taken after the activated carbon filter problems, indicated that the filtration did not work anymore. Another backwash for the carbon filter was made.

The next sample analyses showed that the level of organic matter was staying unchanged. It was concluded that the activated carbon filter did not work as planned after the problems arise with it. Nevertheless, the nanofiltration was kept running till the end of the test runs.

3.2.3 Summary of the test runs

A chart based on the flows in the system is presented in Figure 7 to summarize both of the test runs. Chart shows the flows inside the filtration unit in m³/h. The main events such as cleanings are marked down in to the chart.

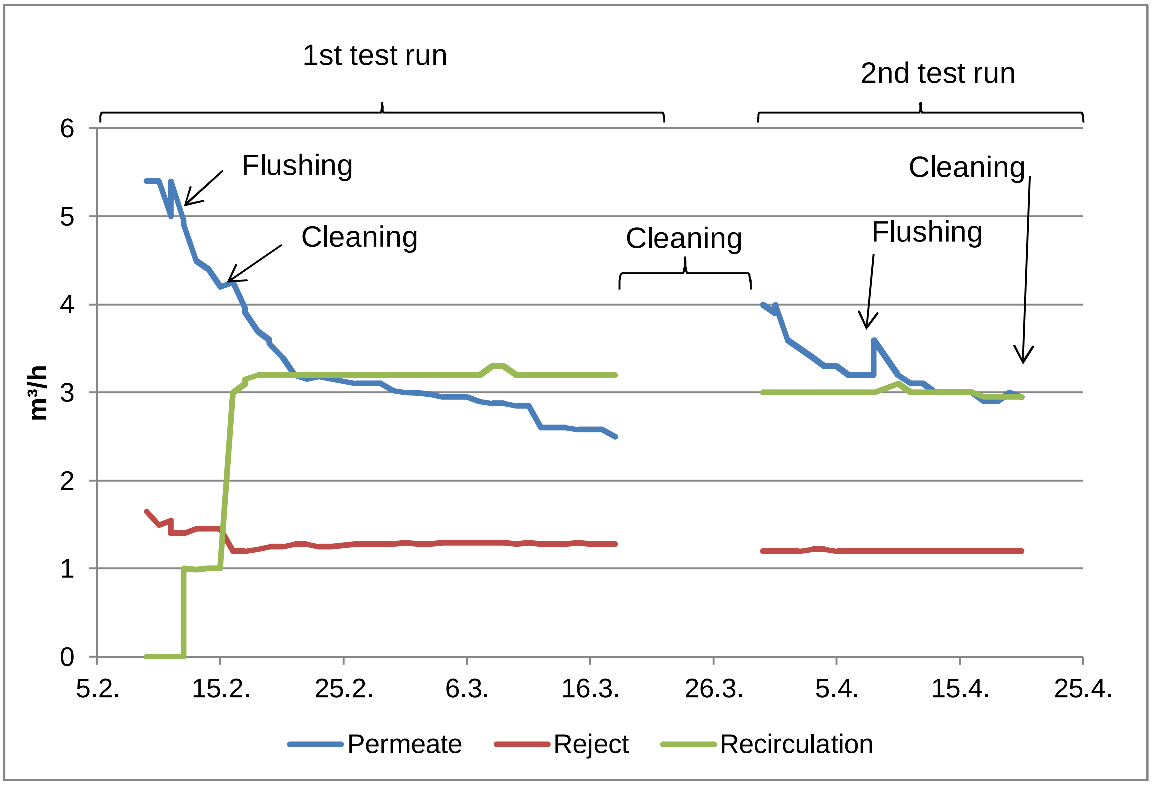


Figure 7. Chart showing the flows inside the filtration system.

Blue line represents the flow of permeate, red represents reject and green is the recirculation that was added after the start. On the left side is seen the first test run and on the right side is the second. The time gap between the test runs are the few days when filtration was turned off. During this time several acid, base and commercial wash solution washes were executed for the membranes.

The first cleaning was executed on 15.02.2016. This is pointed out in the chart. The permeate flow after the wash did not seem to increase much after the first cleaning of the membranes. This indicated that the washing did not remove much scale from the membranes. Additionally, the recirculation of the permeate was changed into maximum. The peak in the beginning of the test was due to the few changes that were made for the process. The reject flow was adjusted and bigger permeate pipe was changed to the system.

The peak half way in the second test run is when the membranes were flushed after carbon got loose from the activated carbon filter. The carbon filter was backwashed already week before when the carbon got loose. Before the peak the carbon filter was backwashed again and the membranes were flushed. This was done in case if some carbon would have got all the way to the membranes.

4 The results

In this part the results of the test runs are discussed in more detailed. The laboratory analyses of the water samples are presented and analysed. The laboratory analyses of the water samples are presented in Appendix 5. The operational journal was already presented in section 3.2 and the journal can be found in Appendix 4. It has to be kept in mind that samples were taken only three times per week and that they represent the situation at that one specific moment. Even though samples were taken regularly and often, they do not provide a full account of the process.

4.1 Water qualities

The samples taken from different stages of the process, helped to monitor the quality of the water and changes in the process. Parameters monitored and analysed during the test were mostly parameters that are mentioned in the drinking water quality regulations of The Ministry of Social Affairs and Health. Ministry has set quality recommendations and requirements for several parameters. Microbial and chemical parameters have requirements and limit values. Water hardness is not mentioned in the requirement or in the recommendations separately. The regulation only states that the water should not be corrosive or create precipitations. [9]

4.1.1 Feed water

The results of the sample analyses proves that the current existing process removes iron and manganese well. Over 96 % of both of these were removed from ground water in the current process. The hardness ions were removed only few percentages from the water. This was known since the existing process is not designed to remove the hardness ions. After the existing process, the iron and manganese levels are well under the limits that Ministry of Social Affairs and Health has set in their drinking water regulations. [9]

The membrane has only few quality parameters that the feed water needs to fulfil. The parameters and the limit values are presented erlier in Table 7. The permanganate value of the feed water was in the upper limit of the parameters during the test. The feed water fulfilled the other requirements. The permanganate value for membranes can be 10 mg/l at the maximum. Figure 8 below shows the permanganate value of the prefiltrated water during the test runs.

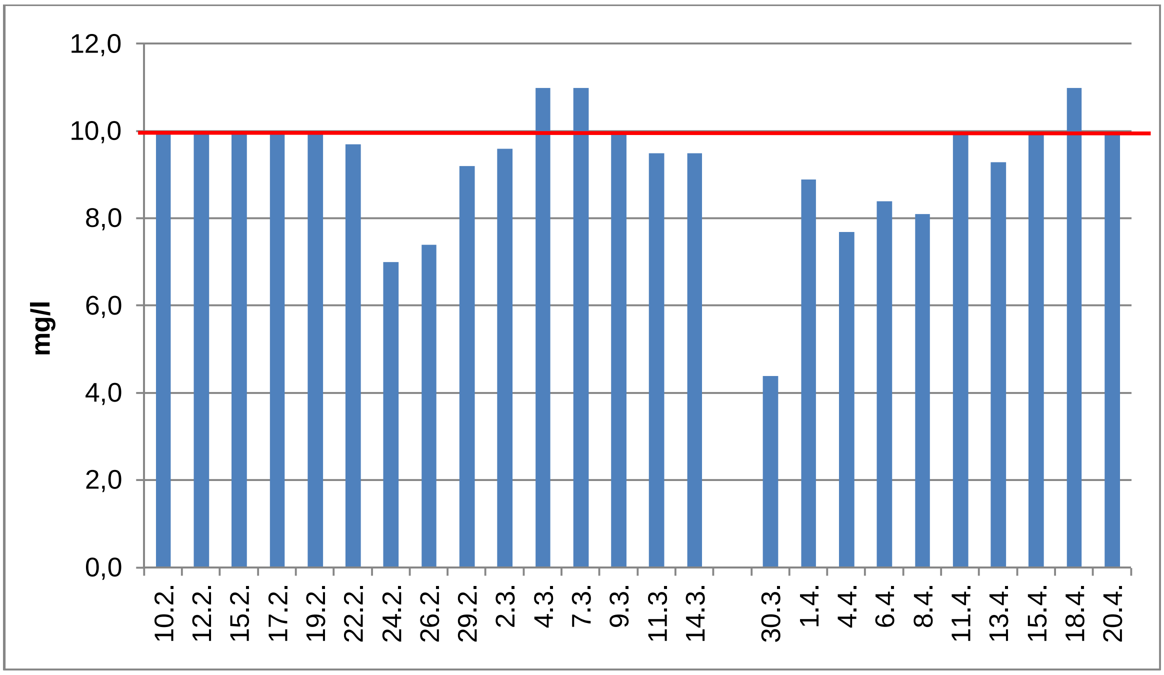


Figure 8. The permanganate value of the feed water and the upper limit for it.

Figure 8 presents the analysed permanganate values and how it fluctuates. The red line presents the maximum limit of permanganate value at 10 mg/l. The feed water permanganate value was varying between 7.4-11 mg/l. When the activated carbon filter was added, the permanganate value was significantly lower in one water sample. This was before the problems with the activated carbon filter occurred.

4.1.2 Permeate

The quality of the permeate was good throughout the test. In the test the aim for the permeate hardness was 1 °dH. The average permeate hardness was 0.08 °dH. This means that the permeate is extremely soft and the permeate harness needs to be increased before letting it to the network. Too soft water is corrosive for the pipes. The hardness is increased with blending the nanofiltered water with water that is unfiltered. This allows the adjustment of the total hardness of the water, that is supplied to the consumers.

The hardness levels during the test are presented in the Figure 9. The aimed 1 °dH hardness is the red line. The green line represents the permeate hardness. Throughout the test the hardness rejection was good and the permeate hardness stayed under the aimed limit. Blue line shows the hardness of the water, that was fed into the membranes.

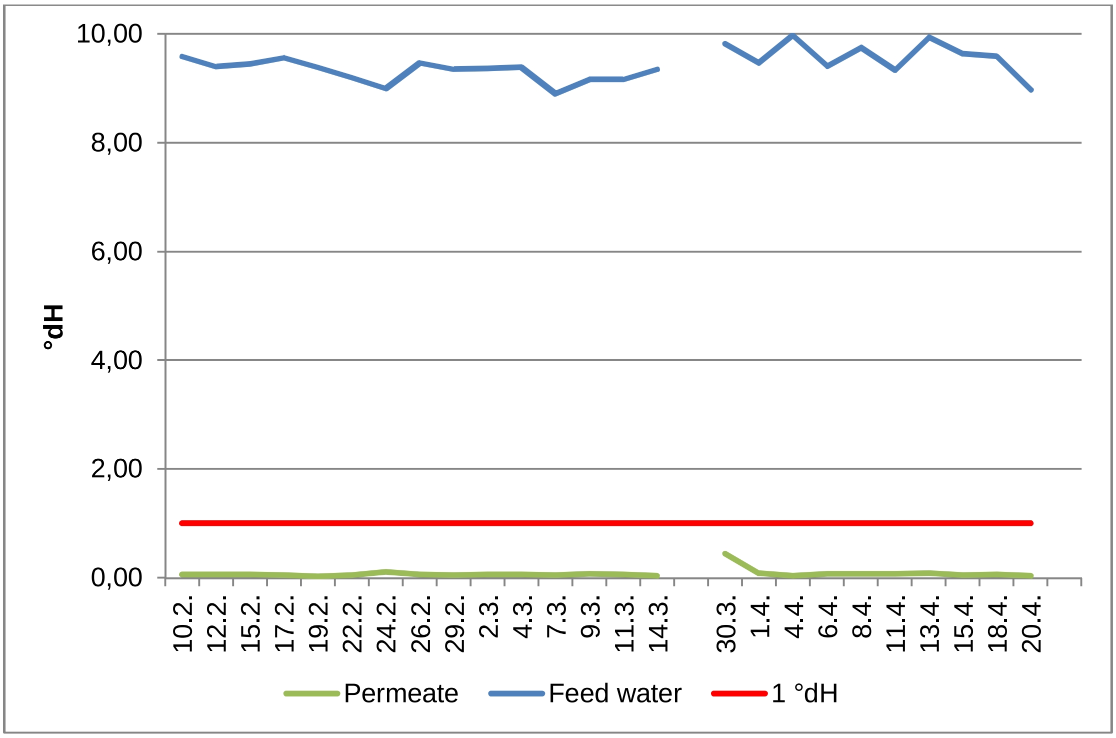


Figure 9. The hardness of the feed water and the permeate. 1 °dH was the goal level of the permeate hardness.

The pH value of the permeate was fluctuating between 5.8 - 8 pH. The lower end of the pH is slightly under the guideline value. The mixing of permeate and non-nanofiltered water will help to maintain the pH value of the guideline. Acidic water is also corrosive for the pipelines.

The microbial quality of the water met the regulation values. The drinking water should not contain E.coli or coliform bacteria. The water taken from the reservoir did not contain any of these. This level stayed throughout the whole process in the plant.

Permeate fulfilled the guideline values set for the drinking water. The feed water to the nanofiltration was already fulfilling the requirements since it is water that is supplied to the consumers. The wish was to lower the hardness and for this purpose nanofiltration has fulfilled its expectations.

Furthermore, nanofiltration lowered the amount of iron and manganese in the permeate. The rejection rate for iron was 65 % on average. The membranes removed about

95 % of manganese. In the permeate sample analyses' the iron and manganese concentration were mostly under the detection limit.

4.1.3 Reject

During the test run the reject was released into the sewers. Meri-Lapin Vesi has permission from the Lapland's Centre for Economic Development, Transport and Environment to lead the reject into the nature, if the full scale nanofiltration would be built in to the plant. The concentrations of each individual analysed parameter (from Table 9) were on average higher in the reject than in feed water.

4.2 Performance of the membrane

The membrane performance can be measured and expressed with several parameters. Monitoring the pressures, flows and water qualities can the membrane performance be determined. Rejection, recovery and flux describe different performances. All of them are important figures on reviewing the membrane performance.

4.2.1 Pressure

The nanofiltration configuration had two pumps. The first pump was pressure increase pump for the feed water. The second pump was integrated into the filtration rig. The first pump increased the pressure up to 6 bar. The pressure then dropped slightly after the activated carbon filter and bag filter. The test rig's pump increased the feed water pressure up to about 18 bar. Figure 10 shows how the pressure meters were located in the process.

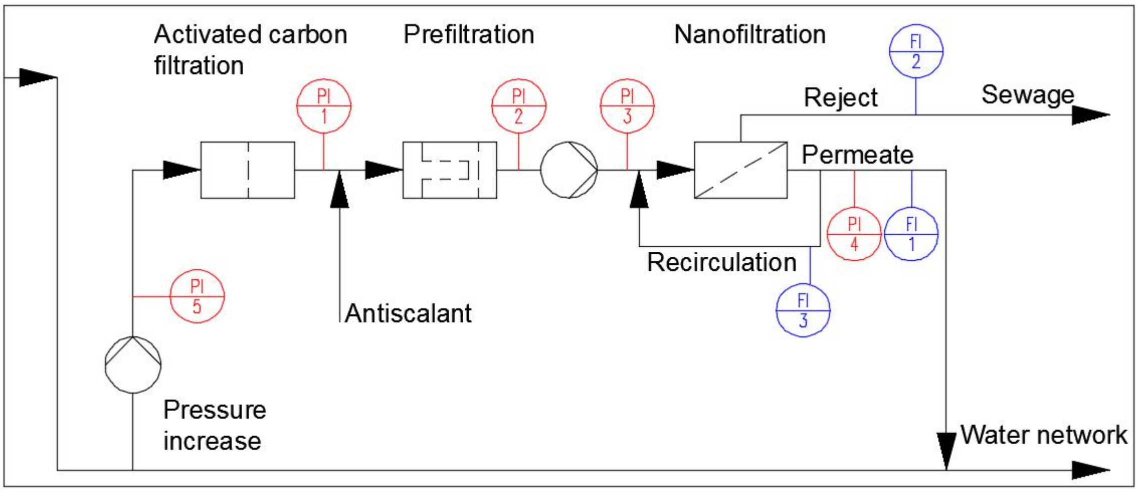


Figure 10. The process flow and instrumentation diagram of the nanofiltration.

All in all 5 pressure meters were monitoring system pressures. Three flow meters were monitoring the out coming flows; permeate, reject and recirculation. Flows of the system were presented in Figure 7. The Figure 11 shows the pressure readings during both of the test runs.

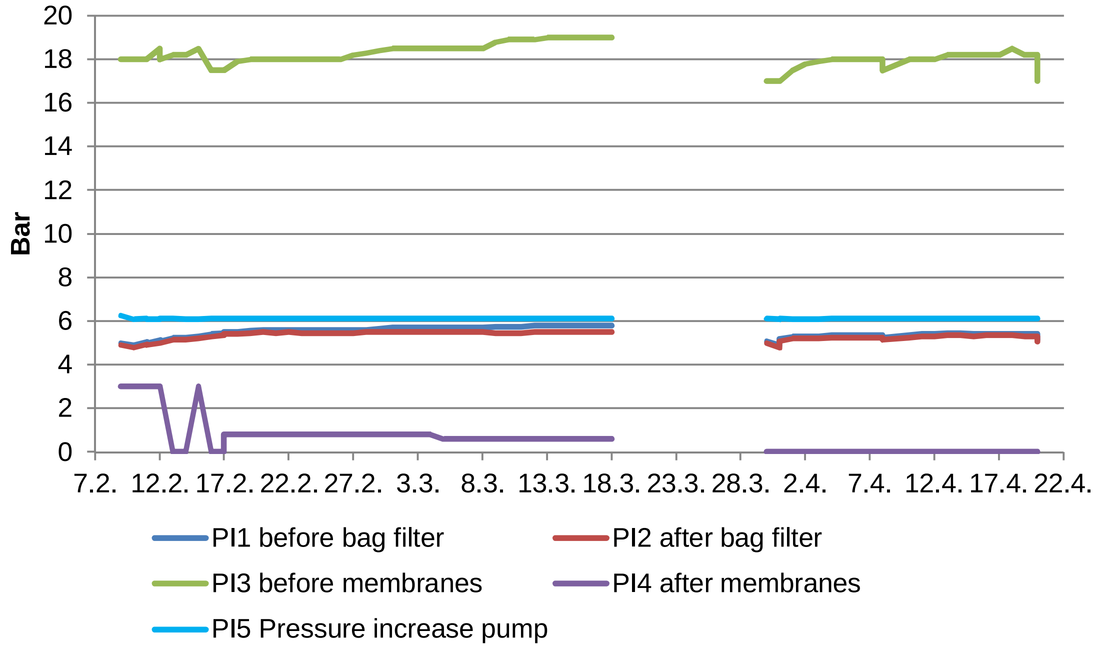


Figure 11. Nanofiltration system pressures during the test run.

Typical operating pressure for nanofiltration and reverse osmosis is 3.4-10.3 bar. [10, p.15.30] Many of the studies mentioned before in section 2.4, operated the nanofiltration in this pressure range. During this test run the nanofiltration was operated with pressure of 18 bar. This pressure is in the membrane's operating limits according to the membrane manufacturers recommendations. However, this feed pressure was higher than the typical nanofiltration operational pressure. High pressure can be a reason why membranes were fouling. When the pressure is over the critical pressure, fouling is stronger.

Manufacturer has also specified a maximum pressure drop over the membrane to be 1 bar. During the test the pressure was measured before the membrane and after the membrane from the permeate line. These two pressures are presented as green and purple lines in the chart. The pressure before membranes was during the test runs about 18 bar. The permeate pipe pressure was extremely low and most of the time under 1 bar. This was considered as oddly high pressure drop for the membranes. A plausible explanation for such a low pressure reading might be faulty pressure meter.

Other explanation might have been that the membranes were fouling extremely strongly causing the high pressure drop. Furthermore, broken membranes were considered as one reason for the pressure drop. It is difficult to say the sole cause for the high pressure drop.

4.2.2 Rejection

The rejection rate for the calcium and magnesium is calculated with the rejection rate formula presented in section 2.2.2. Both calcium and magnesium rejection rates were extremely high during the test runs. For the first test run the average rejection rate for both were over 99%. The rejection rates during the test runs are presented in Figure 12. This high rejection rate was expected with this membrane.

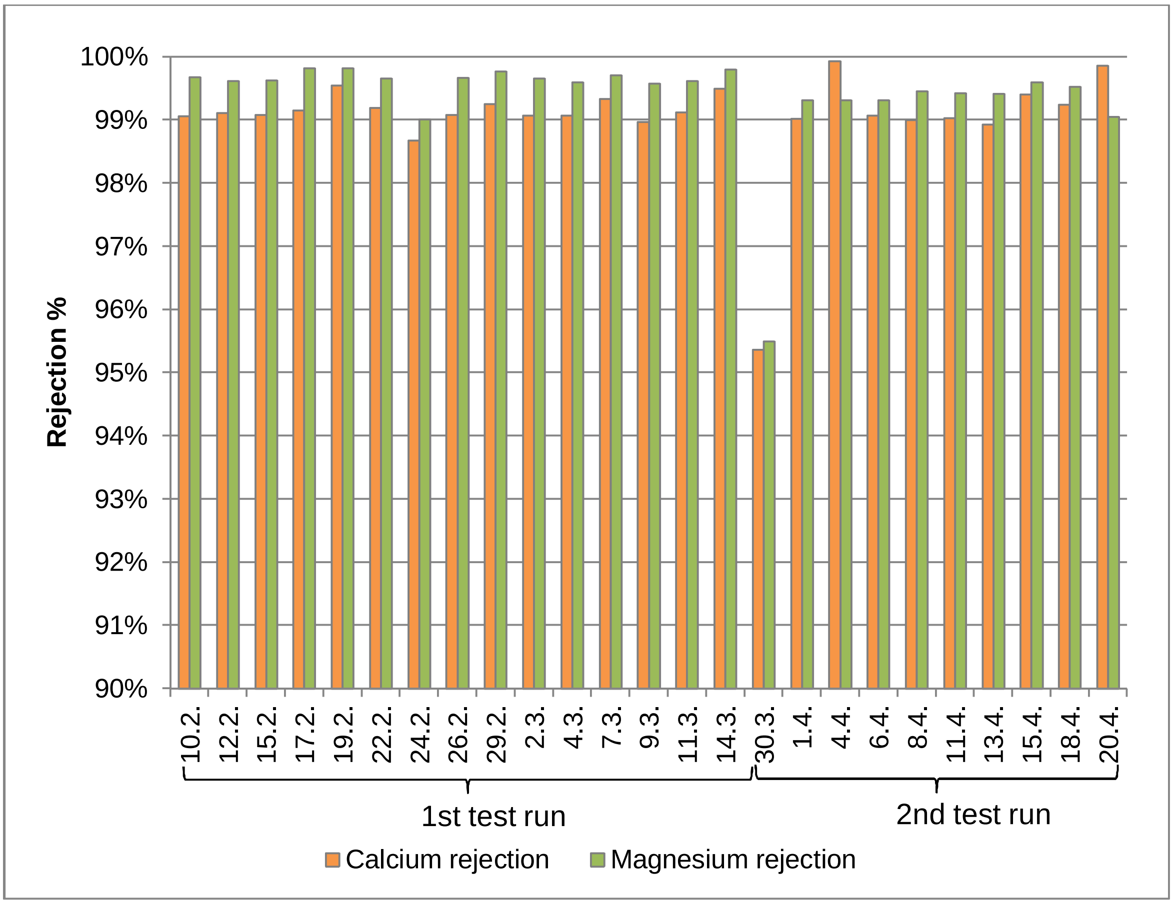


Figure 12. Nanomembrane's rejection rates of the test runs for calcium and magnesium.

With high rejection, much lower permeate hardness than aimed for, was achieved. The average permeate hardness during the test run was 0.08 °dH, when the aim was 1 °dH. Even when the rejection was lower (only about 95 %) on 30th of March, the permeate hardness was under the aimed limit value.

The rejection of iron and manganese were also high. Iron and manganese removal was not the goal with the nanofiltration. Nanomembranes have rejected over 60 % of the iron and over 90 % of the manganese during the first test run. During the second test run the rejection rate was lower than during the first test. The iron and manganese rejections are presented in the Figure 13.

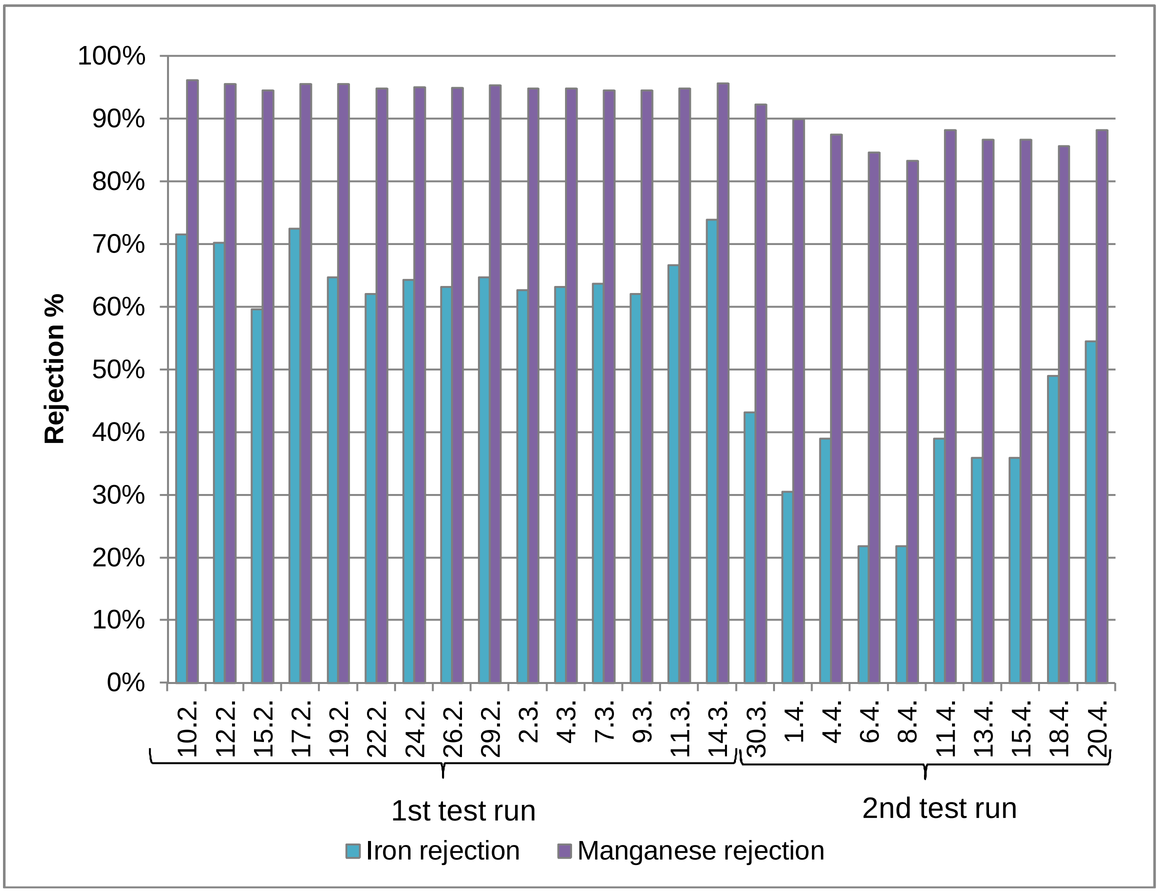


Figure 13. Nanomebrane's rejection rates for iron and manganese.

From the Figure 13 it can be seen that the rejection rate of iron and manganese during the second test were lower than during the first test. This was especially the case with iron. The activated carbon filter removed iron and manganese during the second test and might have caused some of the fluctuation in the rejection.

The absolute values of the iron and manganese were extremely small. The absolute rejections were about 4-7 µg/l during the first test and during the second test about 0.3-2 µg/l. The absolute values during the second test have been so low that the fluctuation can also go in to the analysis measurement error margin.

4.2.3 Recovery

The membranes were aimed to drive with 80 % recovery. This was not achieved at any point of the test. The average recoveries for the test runs were 72 % for the first test and 73 % for the second test. All in all the recovery varied between 66-79 %.

The recirculation was added to the process to increase the recovery. The recirculation adds pressure to the membranes. Higher permeate production is expected when higher pressure is applied. The desired outcome was not achieved, as it can be seen from the Table 7. The permeate production did not increase when recirculation was added. The recovery of the system was declining with a same rate as the permeate flow.

4.2.4 Flux

Usually manufacturers or suppliers give optimum flux for the membranes. After the previous test runs BWT gave a flux of 33.64 l/m²h for the membranes. Later the flux was specified to be 20-22 l/m²h. Typical nanofiltration flux for groundwater is between 22-27 l/m²h [10, p.15.33]. If membranes are operated with too high flux, the membranes will collect a thick particle cake layer on the surface. The cake layer remains thin when the membrane is operated with an optimum flux. This increases the life of the membrane and keeps the permeate production stable. [17] Figure 14 shows the flux of the system during the tests.

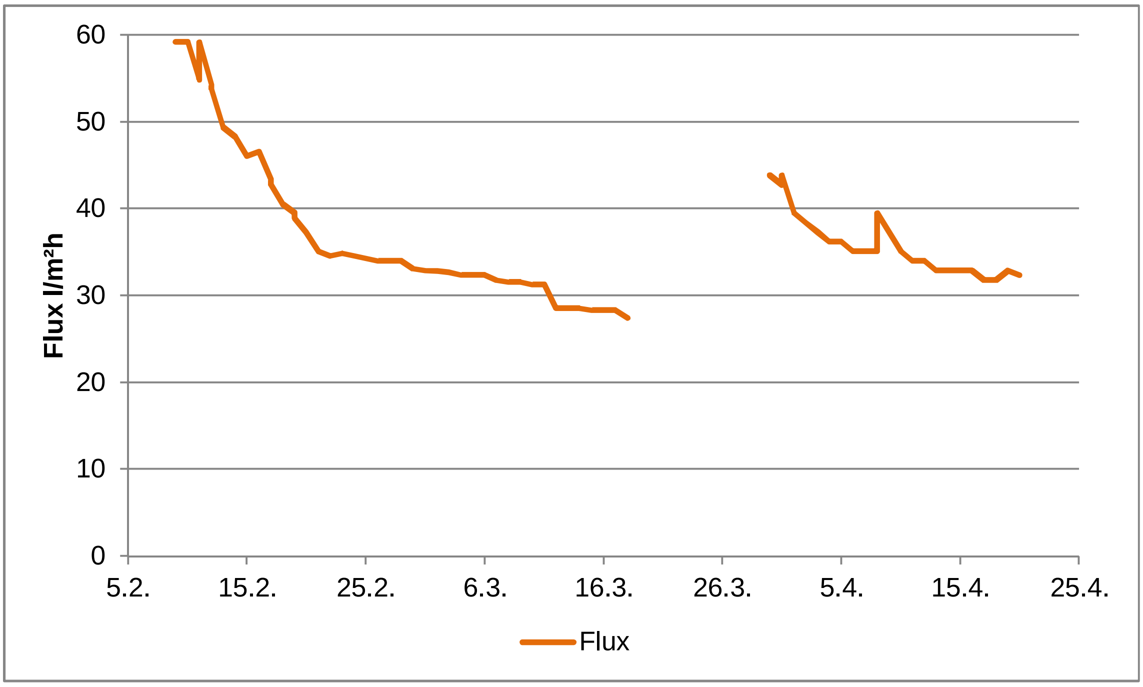


Figure 14. The flux of nanofiltration during the test runs.

In the beginning of the first test the membranes were operated with extremely high flux. The flux decreased rapidly during the first two weeks of the first test run. After that the decrease slowed down. In the second test the decrease was continuous, but it is difficult to say on what level the flux would have settled, since the test was short.

4.3 Bag filter performance

The bag filter was used for removing bigger particles and possible iron and manganese from the feed water. This was done to protect the membranes from fouling. It seems that the bag filter has not had much effect on the feed water quality. Bag filters have had only small rejection rate for the particles. On average the filtration rate was 14-16 % for iron and manganese. In total the bag filter filtrated 5-37 % of the iron and manganese from the feed water. In previous test runs cartridge filters were used and its rejection rate for iron and manganese was said to be 10-50 %. Figure 15 shows the iron and manganese rejection of the bag filter during the test runs.

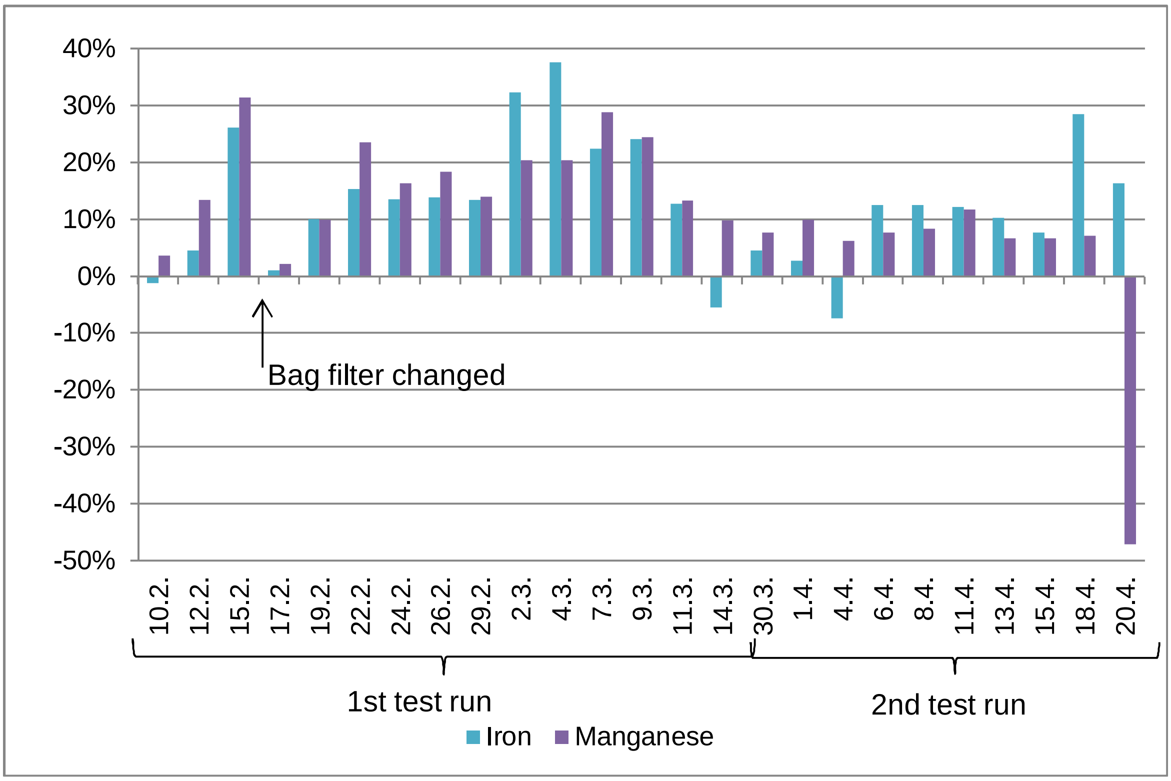


Figure 15. Iron and manganese rejection rates of bag filter during the test runs.

In Figure 15 the negative percentages indicate that particles have got loose from the bag filter. The rejection has fluctuated a lot during the test runs. The efficiency of the bag filter increases when the particles are forming a layer inside it. This can be seen from the beginning of the first test run. When the bag filter was changed to the new one, the rejection dropped until new particle layer was formed inside the filter.

The high percentage for manganese in the last sample seems extremely high, indicating that manganese was getting loose from the bag filter. The absolute amount of manganese was in micro grams. The concentration increase in the analysed water was

0.8 µg/l. Even though the change in percentage was as high as 45 %, it can be noted that the absolute value was small.

4.4 Activated carbon filter performance

Activated carbon filter was added to the process to remove organic matter from the water. Unfortunately this was not succeeded in high amounts. The rejection rates of TOC and permanganate values are presented in Figure 16. Total organic carbon was analysed from both waters only three times during the second test run. TOC values were analysed from activated carbon filtrated water every time. However, from the feed water TOC was not analysed every time and because of this, only three samples can be compared with each other.

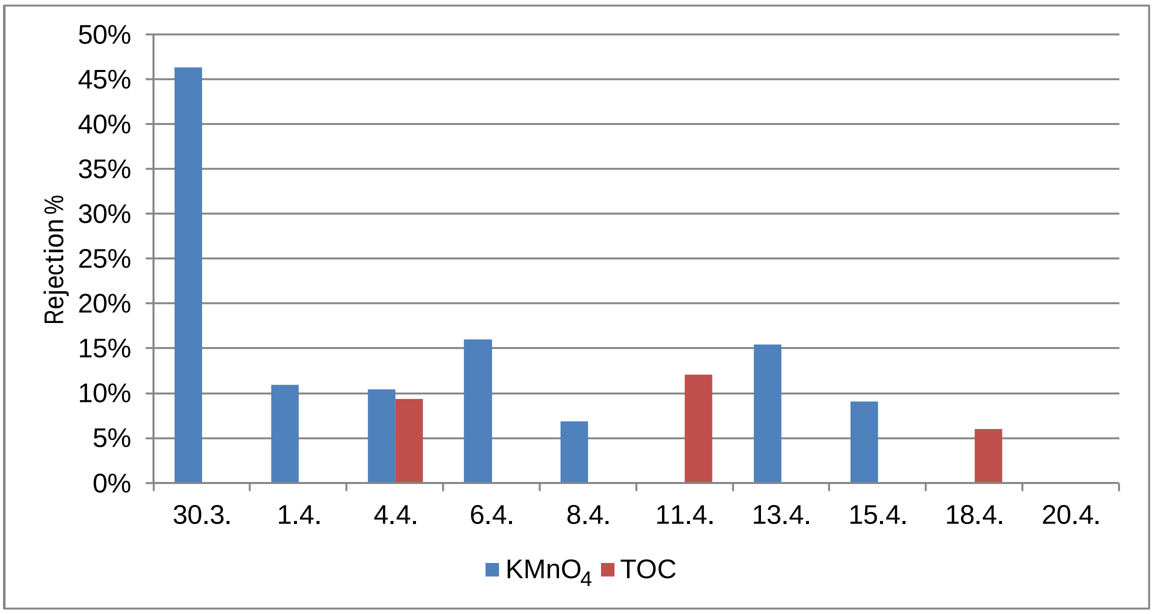


Figure 16. Rejection rates of activated carbon filter for TOC and $KMnO_4$.

The permanganate value of the water was significantly lower when the second test was started. Permanganate value of the water is presented in Figure 8 in page 33. The activated carbon filter removed organic matter in small amounts. In three of the samples, there were not any changes in the permanganate values. These days were 11th, 18th and 20th of April.

On the other hand the activated carbon filter has removed high amounts of iron and manganese from the feed water. The rejection for iron and manganese varied between 30-80 %. From Figure 17 can be seen activated carbon filter rejection rates of iron and manganese. The concentration of iron and manganese of the water that was fed to the membranes were much lower during the second test.

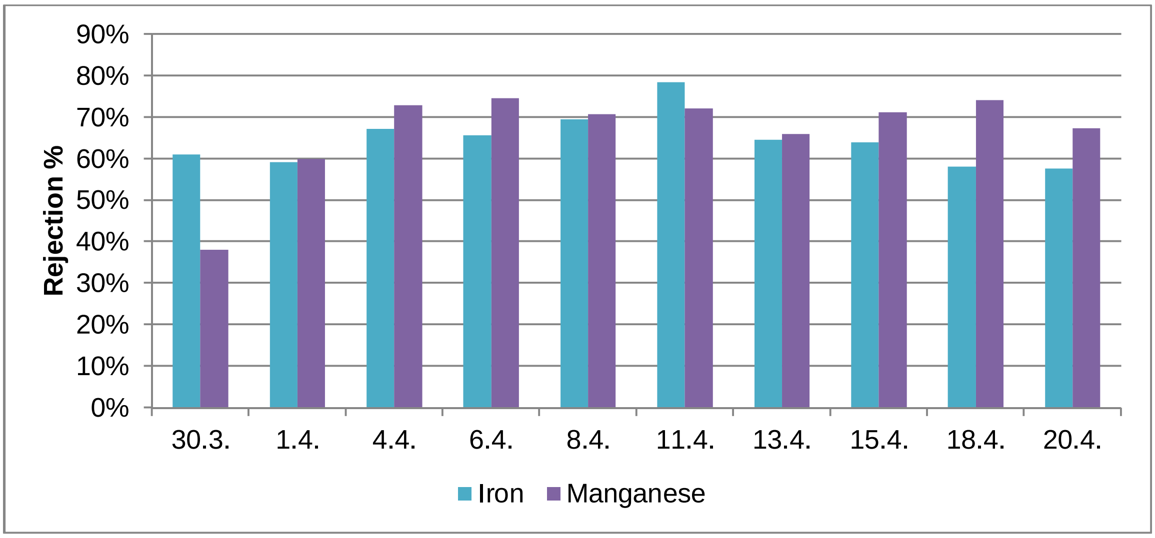


Figure 17. Activated carbon filter rejection rates for iron and manganese.

Activated carbon filter should not remove iron and manganese from the water. Iron and manganese might have been attached to the organic matter. When activated carbon filter was removing organic matter some iron and manganese was removed. For summary, Figure 18 and Figure 19 illustrate the change in the iron and manganese concentration in each water sample before the nanomembranes. The effect of the bag filter seemed to be small for the concentration of iron and manganese. When the activated carbon filter was added, the concentrations of iron and manganese dropped.

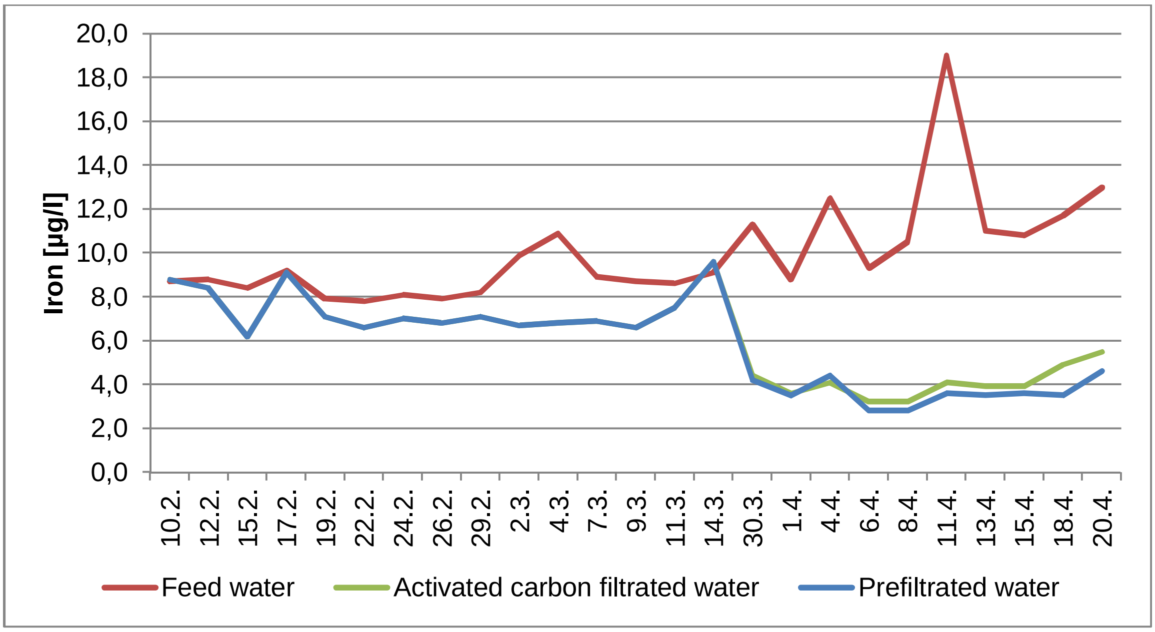


Figure 18. Iron concentration in the waters during the test runs.

The concentration of feed water is presented as red line. Green line presents the concentration after the activated carbon filter. The blue line presents the concentration after the prefilter (bag filter). The effect of the activated carbon filter is seen as a drop in the concentration in the green and blue lines.

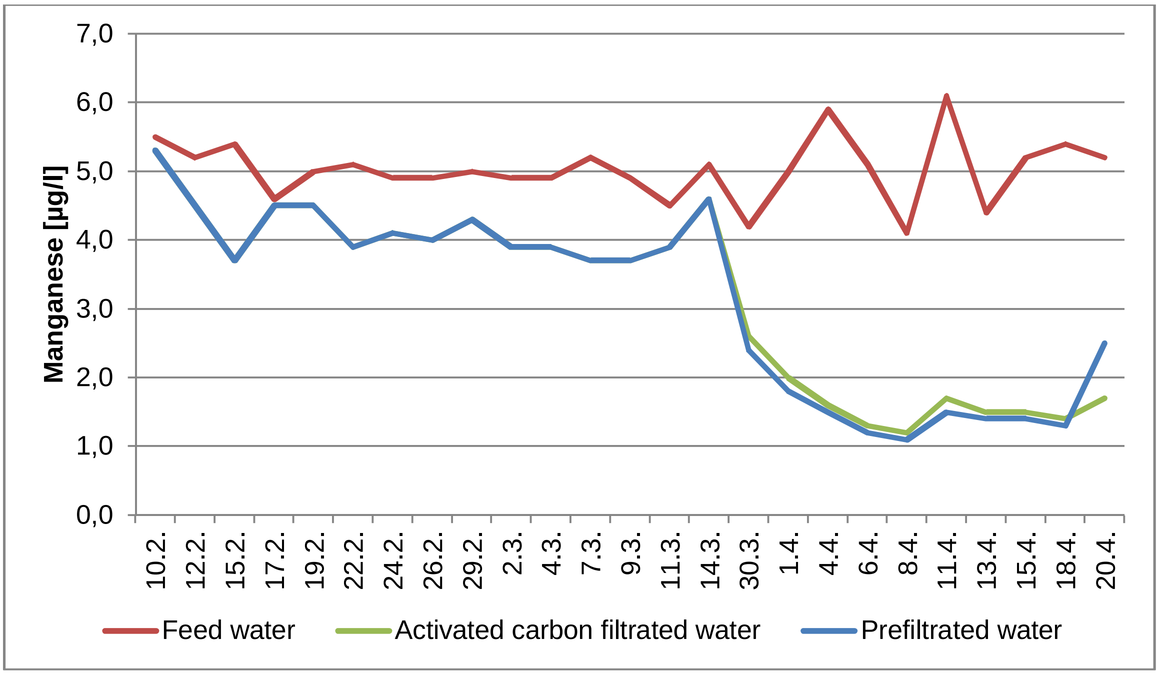


Figure 19. Manganese concentration in the waters during the test runs.

The effect of the activated carbon filter was the same for manganese as for the iron. A drop is seen in the concentration after the activated carbon filter was added. In percentages the concentration drop is high but again looking the absolute values the changes are in µg/l.

4.5 Fouling of the membranes

The membranes started to foul noticeably when the test was started. First it was thought that the malfunction of the antiscalant feed had affected to the fouling. Later it was discussed that organic matter might have caused the fouling.

Closer mass balance study of the reject can reveal the particles which were piling up on the membrane. The analysed concentrations of the reject can be compared to the theoretical values of the reject concentrations. Particles that were fouling on the membranes can be determined based on the concentration difference.

The theoretical reject concentration can be determined by following way: from the feed water and permeate concentrations, and water flows can be calculated the amount of particles in the water. The permeate result will then be deducted from the feed water result. In theory this is then the amount of particles in the reject. This can be then converted to reject concentration. The difference in the actual reject concentration and theoretical concentration can reveal the scaled particles.

The nanofiltration was removing high amounts of calcium and magnesium from the feed water. It is important to study if these particles scaled on the membrane. Also iron and manganese removal was high in the filtration process. Comparing the theoretical and actual concentration, it was noted that these four were scaling on the membrane.

In order to compare the concentration difference of each component, a fouling percentage was calculated. The percentage describes the relationship between the actual analysed concentration and the theoretical concentration of the component. In Figure 20 is presented the percentage of how much from the theoretical concentration scaled on the membrane during the first test run. Negative values represents situation when the actual analysed concentration was higher than theoretical. This means that some of the already scaled particles got loose from the membranes. When talking about the actual concentrations in water it has to be remembered that the amounts are expressed in mg/l for calcium and magnesium, and µg/l for iron and manganese. The changes can be small, such as 0.8 µg/l, but in percentages this can be 45 % change in concentration.

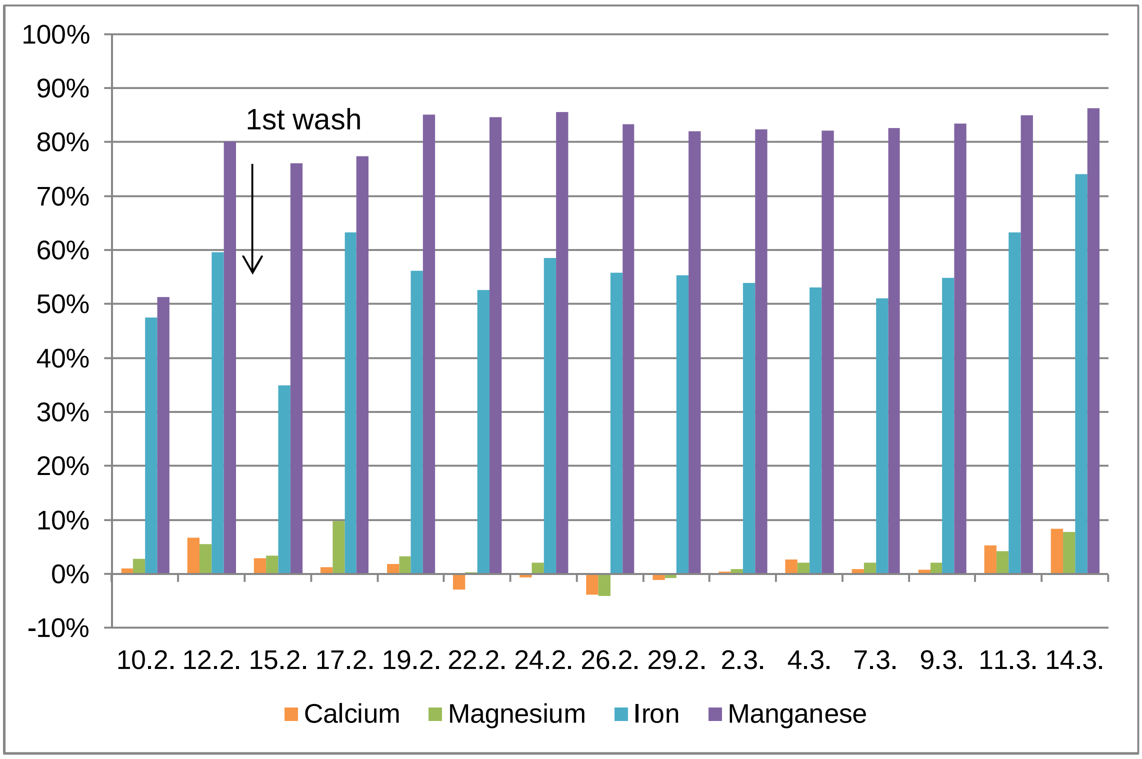


Figure 20. The actual vs. theoretical concentration difference. The percentage how much particles from the theoretical concentration has fouled on the membranes during the first test run.

Based on the results presented in the Figure 20, the scaling of calcium and magnesium were minimal during the first test run. Less than 10 % of the rejected particles scaled on the membrane. Scaling of the iron and manganese in the other hand was high. About 50 % of the iron scaled on the membrane and the scaling percentage was even up to 70 %. For manganese the scaling was almost all the time over 70 % or 80 % during the first test run.

During the second test run the results were similar. The results of theoretical and actual measured concentration difference are presented in Figure 21. The fouling of iron and manganese were on same level as during the first test run. The negative values of calcium and magnesium indicates that already fouled particles have got loose from the membranes. But the absolute values of the calcium and magnesium were so small that the negative values might also fit in the error margin of the analyses.

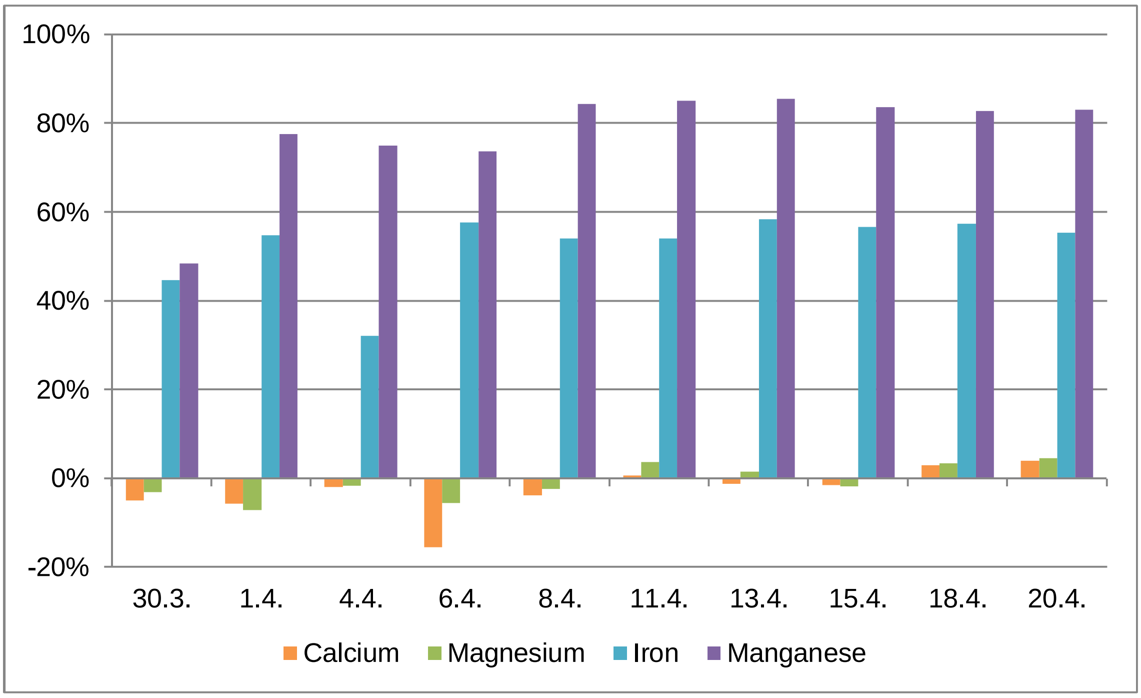


Figure 21. The percentage how much particles from theoretical concentration has fouled on the membranes during the second test run.

From the first wash water sample was analysed high amount of iron and manganese. Calcium and magnesium was also detected from wash water. Unfortunately magnesium and manganese was not analysed from all the samples of the first wash. Detecting these four compounds from wash water goes hand in hand with the concentration difference comparison. The amount of iron and manganese was even higher in the second wash waters. Also calcium and magnesium concentrations were higher in the second wash. The first wash was done week after the filtration was started. The absolute concentrations detected from the wash water samples are presented in Figure 22.

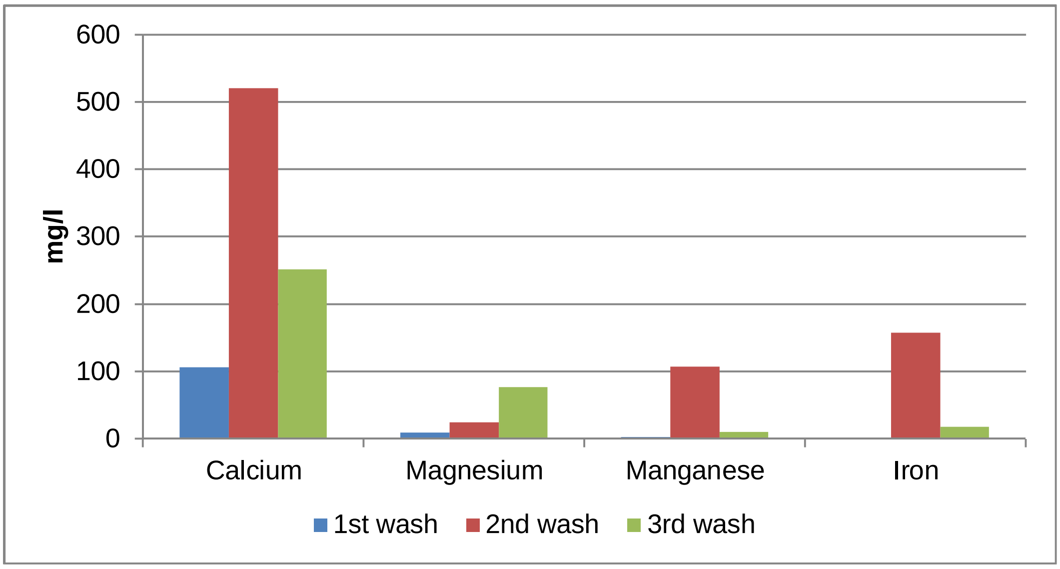


Figure 22. Absolute detected concentrations in wash water samples.

The second wash was done at the end of the first test run, five weeks later than the first wash. The longer run time of the filtration explains why the amounts detected are higher. Also stronger wash chemical was used for cleaning the membranes. When the concentrations in the wash water are compared with the time that filtration was running, the scaling of calcium, magnesium and manganese were similar during both halves of the first test run. The scaling of iron was much higher when compared the wash water to time. Also the third wash after the second run follows this trend. The time related wash water concentrations are presented in Figure 23.

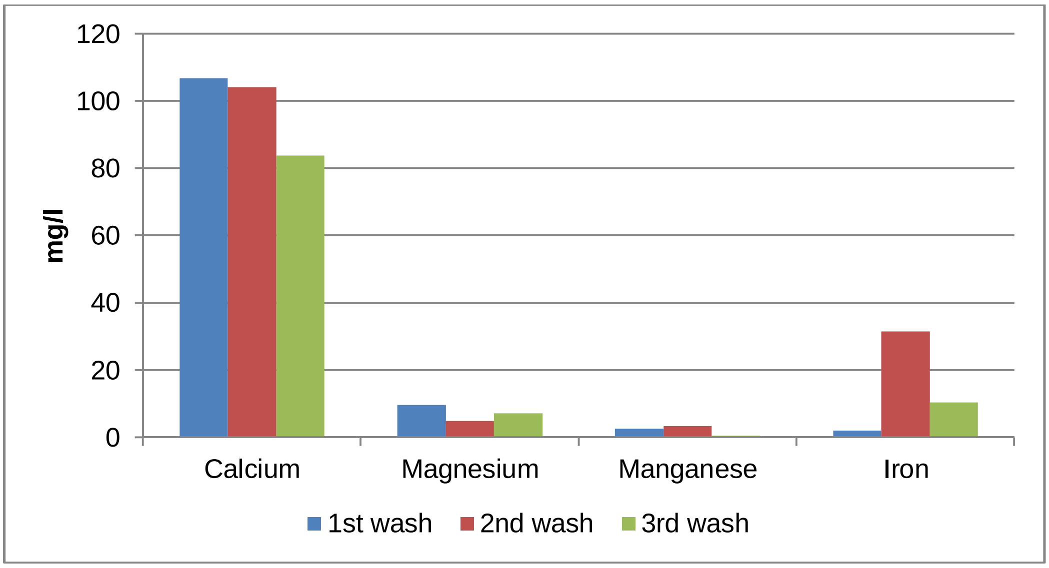


Figure 23. Detected wash water concentrations that are scaled with run time of the tests.

Similar examination of comparing theoretical and analysed reject concentrations is done for the organic matter. This is done by comparing TOC and $KMnO_4$ values in reject. TOC was not analysed from all of the samples each time samples were taken. The results of this comparison are presented in Figure 24.

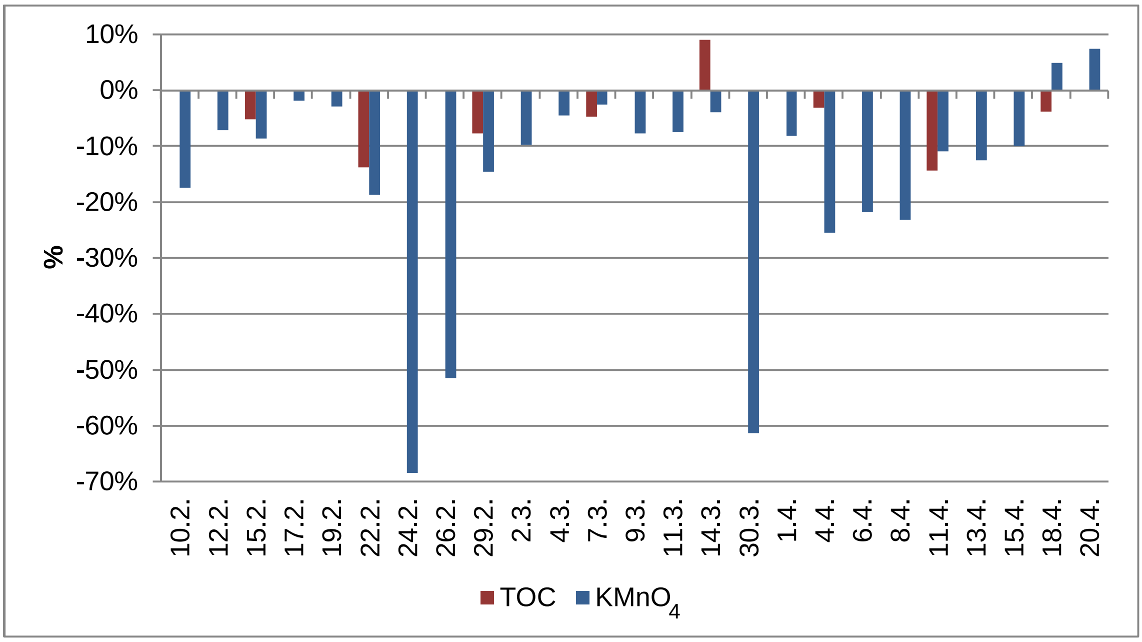


Figure 24. The percentage of how much particles have fouled on the membranes.

Based on Figure 24 organic matter did not scale on the membranes. The negative value of the percentage tells that the actual reject concentration was bigger than theoretical value calculated for the reject. The amount of organic matter was constantly higher in the reject than the theoretical value. But when the wash waters were analysed organic matter was detected from them. From all three wash times organic matter was detected from the samples and especially high amounts from the second wash water samples. The second wash was done with the commercial wash solution. This indicates that the wash solution was good for removing organic matter form membranes.

In another study made in Finland [16] it was concluded that the TOC did not have significant role in the fouling of the membranes. The chemical process and sand filtration before the nanofiltration was held more responsible of the fouling of the membrane. Antiscalant was found to add fouling of the membranes in one of the plants where nanofiltration was tested. [16] This shows how case sensitive nanofiltration membrane fouling is. More exact fouling factors can be determined if the membranes are opened and analysed. This was not done for the membranes used in Meri-lapin Vesi.

5 Cost calculation

Content of this chapter is confidential and therefore, not published.

6 Conclusion

Meri-lapin Vesi Oy wished to know how nanofiltration suits for their need to lower water hardness and how much the operational costs of the nanofiltration system would be. Test runs with nanofiltration were executed in their water treatment plant in spring 2016. Two different tests were executed during the test period. The tests were conducted with DOW Filmtech membrane NF90-4040.

The scaling of the membranes was high in the tests and the target of permeate production was not achieved. The recovery of the membrane was on average 72 %, when the aim was 80 %. The operational flux during the test was over the critical flux. This might have been contributing the fouling of the membranes. Organic matter was considered as one of the causes of membrane fouling. Calcium, magnesium, iron and manganese were all found to form scale on the membranes. However, the real reasons behind the fouling are difficult to state certainly. In order to confirm this assumption, the membranes should have been opened and examined more closely to find out the full composition of the scaled substances.

Nanofiltration was noted to be suitable for removing hardness from water. The expected removal rates were exceeded. Nanofiltration removed over 99 % of the hardness ions: calcium and magnesium. Large amounts of iron and manganese were also noted to be removed by nanofiltration.

Some of the operational parameters during the test were considered too high. These parameters should be revised when designing a full size nanofiltration plant for Meri-Lapin Vesi. The flux and pressure of the nanofiltration during the test were on the higher side. By lowering these two parameters, the scaling of the membranes would most likely decrease.

The cost evaluation reveals that the operational costs of the nanofiltration in Meri-Lapin Vesi would fit in the price range the other nanofiltration studies gave. Nevertheless, all of these studies cannot be compared with each other because of the length of the time span during which they have been conducted. Also differences in national levels cannot be compared. The comparison of the prices between each other is more indicative.

This gives the idea what kind of results could be expected. The operational costs for Meri-Lapin Vesi were calculated based on the results of the tests. Some operational parameters were changed in the calculations to correspond the results because the original aims were not achieved in the tests.

References

1 Korkka-Niemi K, Salonen VP. Maanalaiset vedet –pohjavesigeologian perusteet. Turun yliopiston täydennyskoulutuskeskus, 1996: 78-79.

2 Gray N.F. Drinking water quality. Second edition. Cambridge University Press; 2008.

3 Lenntech: water treatment sollutions [Internet]. c.1998-2016. Water hardness calculator.[Date unknown] [Cited 2016 February 25] Available from: http://www.lenntech.com/ro/water-hardness.htm

4 Suomen Rakennusinsinöörien Liitto RIL Ry. RIL 124-1 Vesihuolto I. Vammalan Kirjapaino Oy, 2003.

5 Insinööritöimisto Kaiko Oy. Vesikirja. Fifth edition. Amer-yhtymä Oy Weilin+Göös; 1988.

6 HSY, Veden kovuus Suomessa [Internet]. [Cited 2016 March 23] Available from: https://www.hsy.fi/fi/opettajalle/sahkoisetoppimateriaalit/Documents/Ammattia%20oppimassa/Hoitoala/vedenkovuus_suomessa_fiksu.pdf

7 Helsingin yliopisto [Internet]. Kemian opetuksen keskus: Veden kovuus. [Date unkown] [Cited 2016 March 23] Available from: http://www.kemianluokka.fi/files/uudet/Puskurikapasiteetti%20oppilas_Aga.pdf

8 World Health Organization. Guidelines for drinking-water quality [Internet]. Fourth edition. World Health Organization; 2011. [Cited 2016 April 8] Available from: http://apps.who.int/iris/bitstream/10665/44584/1/9789241548151_eng.pdf.

9 Sosiaali- ja terveysministeriön asetus talousveden laatuvaatimuksista ja valvontatutkimuksista, 1352/2015. (November 17, 2015)

10 The American Water Works Association (AWWA), The American Society of Civil Engineers (ASCE). Water Treatment Plant Design. Fifth edition. United States: McGraw-Hill Education ; 2012.

11 Water Treatment Principles and Design. Second edition. United States of America: John Wiley & Sons; 2005.

12 Van der Bruggen B, Everaert K, Wilms D, Vandecasteele C. Application of nanofiltration for removal of pesticides, nitrate and hardness from groundwater: rejection properties and economic evaluation. Journal of membrane Science. 2001; 193: 239-248.

13 Van Der Bruggen B, Vandecasteele C. Removal of pollutants from surface water and groundweater by nanofiltration: overview of possible applications in the drinking water industry. Environmental pollution. 2003; 112: 435-445.

14 Härmä V. Nanosuodatus talousveden valmistuksen viimeistelevänä käsittelynä.[Nanofiltration as a refining unit process in water treatment]. Otaniemi: Teknillinen korkeakoulu; 1996. (Teknillisen korkeakoulun vesihuoltotekniikan laboratorion julkaisu 19).Finnish.

15 Synder filtration [Internet]. Ultrafiltration elements: Spiral wound configuration. [Date unkown] [Cited 2016 February 18] Available from: http://www.synderfiltration.com/wp-content/uploads/2013/03/Spiral-Wound-Configuration-3-8.jpg

16 Härmä V. Nanosuodatus talousveden laadunparantajana suomalaisilla pintavesilaitoksilla.[Nanofiltration for water quality improvement in Finnish surface waterworks]. Otaniemi: Teknillinen korkeakoulu; 1999. (Teknillisen korkeakoulun vesihuoltotekniikan laboratorion julkaisu 22). Finnish.

17 Korhonen K. Mikrosuodatus ja sen soveltuvuus nanosuodatuksen esikäsittelymenetelmäksi. [Microfiltration and its suitability as a pretreatment method for nanofiltration]. Otaniemi: Teknillinen korkeakoulu; 2000. Teknillisen korkeakoulun vesihuoltotekniikan laboratorion julkaisu 27). Finnish.

18 Gaida A, Bablon G, Turner G, Franchet J, Protais JC. Performance of 3 years' operation of nanofiltration plants. Desalination. 1998; 117: 149-158.

19 Nurminen M. Nanosuodatuksen kustannukset talousveden käsittelyssä Suomessa. [Cost of Nanofiltration in Potable Water Production in Finland]. Otaniemi: Teknillinen korkeakoulu; 2001. Teknillisen korkeakoulun vesihuoltotekniikan laboratorion julkaisu 30. Finnish.

20 Bergman R.A. Cost of membrane softening in Florida. American Water Works Association, May 1996: 32-43

21 Gorenflo A, Velázquez-Padrón D, Frimmel FH. Nanofiltration of a German groundwater of high hardness and NOM content: performance and costs. Desalination. 2002; 151: 253-265.

22 US Inflation Calculator [Internet]. c.2008-2015. Inflation calculator. [Cited 2016 April 7] Available from: http://www.usinflationcalculator.com/

23 European Central Bank [Internet]. Euro foreign exchange reference rates: 6 April 2016. c.2016 [Cited 2016 April 7] Available from: https://www.ecb.europa.eu/stats/exchange/eurofxref/html/index.en.html

24 Suomen virallinen tilasto (SVT): Kuluttajahintaindeksi [Internet]. [2016 January 15]. 2015, Rahanarvonkerroin 1860–2015 . Helsinki: Tilastokeskus [Cited 2016 February 10].
Available from: http://www.stat.fi/til/khi/2015/khi_2015_2016-01-15_tau_001.html

25 Liikanen R, Yli-Kuivala J, Tenhunen J, Laukkanen R. Cost and environmental impact of nanofiltration in treating chemically pre-treated surface water. Desalination. 2006; 201: 58-70.

26 Elazhar F, Tahaikt M, Achatei A, Elimidaoui F, Taky M, El Hannouni F, Laaziz I, Jariri S, El Amrani M, Elimidaoui A: Economical evaluation of the fluoride removal by nanofiltration. Desalination. 2009; 249: 154-157.

27 Costa A, de Pinho M: Performance and cost estimation of nanofiltration for surface water treatment in drinking water production. Desalination. 2006; 196: 55-65

Appendix 1. Nanofiltration plants around the world: design parameters and costs

	Unit	Mainz, Germany		Belgium					
Year of the study		2002		2001					
Water source		Groundwater		Groundwater					
Capacity	m^3/d	27000	11000	48000	48000	48000	48000	48000	17500
Membrane		NF200B, FilmTech		NF70 8040, DOW					
Recovery	%	85	85	80	80	80	80	80	80
Operating pressure	bar	5,5	5,5	8	5	10	15	20	8
Removal of		Hardness and NOM		Pesticides, nitrate and hardness					
Rejection rate	%	>74, >95		90-95, 76, 95					
Chemical	€/m³	0,004		0,025					
Energy	€/m³	0,051		0,002					
Work	€/m³								
Concentration	€/m³	0,037		0,019					
Membrane replacement	€/m³	0,022							
Maintenance	€/m³	0,010							
Other	€/m³			0,022					
Amortization	€/m³	0,104							
Cost	€/m³	**0,230**	**0,270**	**0,126**	**0,128**	**0,132**	**0,145**	**0,157**	**0,260**

	Unit	Plantation-Central Plant, USA	Fort Mayers, USA	Collier County	Indian River County South, USA	Dunedin, USA	Boynton Beach, USA	St. Lucie West Development, USA
Year of the study		1996	1996	1996	1996	1996	1996	1996
Water source		Groundwater	Groundwater	Groundwater	Groundwater	Groundwater	Groundwater	Groundwater
Capacity	m^3/d	45400	45400	45400	22700	22700	15100	3800
Membrane		Fluid System	Hydranautics	Hydranautics	Fluid System & Hydranautics	Hydranautics	DOW, Filmtech	Hydranautics
Recovery	%	85	90	90	85	83	85	85
Operating pressure	bar	8,96	10,69	7,58	7,58-8,25	7,24	7,24	6,85
Removal of		Hardness	Hardness	Hardness	Hardness	Hardness	Hardness	Hardness
Rejection rate	%	94	44	69	90	63	78	72
Chemical	€/m³	0,040	0,026	0,026	0,040	0,066	0,013	0,119
Energy	€/m³	0,040	0,053	0,053	0,079	0,066	0,040	0,224
Work	€/m³	0,040	0,066	0,053	0,053	0,053	0,132	0,198
Concentration	€/m³							
Membrane replacement	€/m³	0,026	0,026	0,026	0,013	0,040	0,026	0,040
Maintenance	€/m³							
Other	€/m³	0,013	0,040	0,026	0,013	0,026	0,013	0,132
Amortization	€/m³							
Cost	€/m³	**0,158**	**0,198**	**0,198**	**0,198**	**0,251**	**0,238**	**0,700**

	Unit	Kempele/Tuohino, Finland	Björkby/Mustasaari, Finland	Puntari/Laitila, Finland	Dämman/Espoo, Finland	Valada, Portugal	Marocco	Paris, France
Year of the study		2001	2001	2001	2006	2006	2009	2000
Water source		Groundwater	Groundwater	Groundwater	Surface water	River water	Groundwater	River water
Capacity	**m³/d**	700	240	170	18000	100000	2400	1470
Membrane		NF90-400	NF55-400 & 8040-255, FilmTech	NF255-400	NF255-400	NF200B-400	NF90 8040 FilmTech	NF70 DOW
Recovery	**%**		75	71	83	90	84	85
Operating pressure	**bar**	8,3 -12	3-5	5,7	6	15	10	8
Removal of		Manganese, iron and humus	NOM	Fluoride and aluminium	NOM	NOM and salt	Fluoride	Pesticides and organic matter
Rejection rate	**%**	75 , 98	90	76-95 , 80		>97, 86	97,8	90, 86
Chemical	**€/m³**	0,008	0,023	0,002		0,010	0,050	
Energy	**€/m³**	0,023	0,021	0,019		0,048	0,040	
Work	**€/m³**	0,060	0,023	0,029			0,000	
Concentration	**€/m³**	0,048	0,000	0,000		0,037		
Membrane replacement	**€/m³**	0,123	0,115	0,206		0,017	0,009	
Maintenance	**€/m³**					0,010	0,025	
Other	**€/m³**					0,023	0,062	
Amortization	**€/m³**					0,067	0,026	
Cost	**€/m³**	**0,262**	**0,181**	**0,256**	**0,110**	**0,214**	**0,212**	**0,120**

Appendix 2. Nanofiltration membrane product information sheet

Product Information

DOW

DOW FILMTEC™ Membranes

DOW FILMTEC NF90 Nanofiltration Elements for Commercial Systems

Features

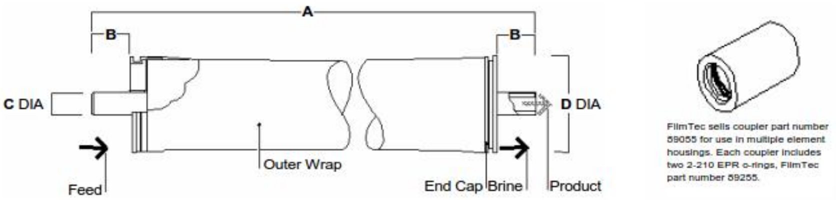
The DOW FILMTEC™ NF90 membrane elements provide high productivity performance while removing a high percentage of salts, nitrate, iron and organic compounds such as pesticides, herbicides and THM precursors. The low net driving pressure of the NF90 membrane allows the removal of these compounds at low operating pressures.

Product Specifications

Product	Part Number	Applied Pressure psig (bar)	Permeate Flow Rate gpd (m^3/d)	Stabilized Salt Rejection (%)
NF90-2540	149982	70 (4.8)	680 (2.6)	>97.0
NF90-4040	149983	70 (4.8)	2,000 (7.6)	>97.0

1. Permeate flow and salt rejection based on the following test conditions: 2,000 ppm $MgSO_4$, 77°F (25°C) and 15% recovery at the pressure specified above.
2. Permeate flows for individual NF90-2540 elements may vary by -20% / +30%. NF90-4040 individual elements may vary -15% / +50%.
3. Developmental products available for sale.

Figure 1



Product	Dimensions – Inches (mm) A	B	C	D
NF90-2540	40.0 (1,016)	1.19 (30)	0.75 (19)	2.4 (61)
NF90-4040	40.0 (1,016)	1.05 (27)	0.75 (19)	3.9 (99)

1. Refer to DOW FILMTEC Design Guidelines for multiple-element systems. 1 inch = 25.4 mm
2. NF90-2540 has a tape outer wrap. NF90-4040 has a fiberglass outer wrap.

Operating Limits

- Membrane Type — Polyamide Thin-Film Composite
- Maximum Operating Temperature — 113°F (45°C)
- Maximum Operating Pressure — 600 psi (41 bar)
- Maximum Feed Flow Rate - 4040 elements — 16 gpm (3.6 m^3/hr)
  - 2540 elements — 6 gpm (1.4 m^3/hr)
- Maximum Pressure Drop - tape wrapped — 13 psig (0.9 bar)
  - fiberglassed — 15 psig (1.0 bar)
- pH Range, Continuous Operation[a] — 2 – 11
- pH Range, Short-Term Cleaning (30 min.)[b] — 1 – 12
- Maximum Feed Silt Density Index — SDI 5
- Free Chlorine Tolerance[c] — <0.1 ppm

[a] Maximum temperature for continuous operation above pH 10 is 95°F (35°C).
[b] Refer to Cleaning Guidelines in specification sheet 609-23010 for NF90.
[c] Under certain conditions, the presence of free chlorine and other oxidizing agents will cause premature membrane failure. Since oxidation damage is not covered under warranty, DOW FILMTEC recommends removing residual free chlorine by pretreatment prior to membrane exposure. Please refer to technical bulletin 609-22010 for more information.

Important Information

Proper start-up of reverse osmosis water treatment systems is essential to prepare the membranes for operating service and to prevent membrane damage due to overfeeding or hydraulic shock. Following the proper start-up sequence also helps ensure that system operating parameters conform to design specifications so that system water quality and productivity goals can be achieved.

Before initiating system start-up procedures, membrane pretreatment, loading of the membrane elements, instrument calibration and other system checks should be completed.

Please refer to the application information literature entitled "Start-Up Sequence" (Form No. 609-02077) for more information.

Operation Guidelines

Avoid any abrupt pressure or cross-flow variations on the spiral elements during start-up, shutdown, cleaning or other sequences to prevent possible membrane damage. During start-up, a gradual change from a standstill to operating state is recommended as follows:

- Feed pressure should be increased gradually over a 30-60 second time frame.
- Cross-flow velocity at set operating point should be achieved gradually over 15-20 seconds.
- Permeate obtained from first hour of operation should be discarded.

General Information

- Keep elements moist at all times after initial wetting.
- If operating limits and guidelines given in this bulletin are not strictly followed, the limited warranty will be null and void.
- To prevent biological growth during prolonged system shutdowns, it is recommended that membrane elements be immersed in a preservative solution.
- The customer is fully responsible for the effects of incompatible chemicals and lubricants on elements.
- Maximum pressure drop across an entire pressure vessel (housing) is 30 psi (2.1 bar).
- Avoid static permeate-side backpressure at all times.

DOW FILMTEC™ Membranes
For more information about DOW FILMTEC membranes, call the Dow Water & Process Solutions business:
North America: 1-800-447-4369
Latin America: (+55) 11-5188-9222
Europe: (+32) 3-450-2240
Pacific: +60 3 7958 3392
Japan: +813 5460 2100
China: +86 21 2301 1000
www.dowwaterandprocess.com

Appendix 3. Filtration test rig product information sheet

RO-2500

BWT - The Water Company

HOH-käänteisosmoosilaitteet on rakennettu ruostumattomasta teräksestä valmistettuun runkoon valmiiksi kokonaisuudeksi.

- Pieni tilantarve
- Yksinkertaiset sähkö- ja LVI-asennukset

Toiminta

Käänteisosmoosi (RO) on kalvoerotustekniikkaa, jossa korkean veden paineen avulla erotetaan raakaveteen liuenneet suolat (ionit) ja päästetään kalvon läpi puhtaat vesimolekyylit. Käänteisosmoosia käytetään pääasiassa silloin, kun suolan poistaminen on tärkeää.

Tarkalleen ottaen käänteisosmoosi on liuenneiden suolojen erottamista vedestä. Kyse ei ole ionien poistamisesta, kuten ioninvaihdossa. Käänteisosmoosilla liuenneet suolat erotetaan vedestä lähes sataprosenttisesti. RO-kalvon huokoset ovat niin pieniä, etteivät niitä läpäise edes mikro-organismit, kuten bakteerit ja pyrogeenit.

Puhdas vesi (permeaatti) voidaan kerätä varastosäiliöön ja pumpata sieltä käyttökohteeseen. Epäpuhtaudet sisältävä vesi (konsentraatti) johdetaan viemäriin.

RO-2500-laitteissa on PLC-ohjaus vakiovarusteena. Ohjauspaneelissa on LCD-näyttö, jolta laitteen toimintaparametrit on helppo lukea. PLC-ohjaus mahdollistaa useita toimintoja, kuten liitännän BUS-järjestelmään ja laitteen käyttöparametrien uudelleen ohjelmoinnin asiakkaan tarpeiden mukaisesti.

Lisävarusteet

- Antiskalantin annosteluyksikö
- CIP-yksikkö

Tekninen tieto	RO-2510	RO-2520	RO-2530	RO-2540	RO-2550
Kapasiteetti, l/h*	2300	2700	3300	4000	5000
Saanto maks., %	80	80	80	80	80
Suolanpoisto n., %	> 98	> 98	> 98	> 98	> 98
Johtokyky n., µS/cm	< 20	< 20	< 20	< 20	< 20
Sähkönsyöttö, V/Hz	3x400/50	3x400/50	3x400/50	3x400/50	3x400/50
Syöttöteho, kW	5,5	5,5	7,5	7,5	7,5
Tuloyhde, Ø"	1½	1½	1½	1½	1½
Lähtöyhde, Ø"	1¼	1¼	1¼	1¼	1¼
Viemäriyhde, Ø"	1	1	1	1	1
Tuloyhteen korkeus, mm	785	785	785	785	785
Lähtöyhteen korkeus, mm	1510	1510	1510	1510	1510
Viemäriyhteen korkeus, mm	1375	1375	1375	1375	1375
Maks. veden lämpötila, °C	25	25	25	25	25
Tulopaine, min./maks., bar	3/7	3/7	3/7	3/7	3/7
Mitat L x S x K, mm	1300 x 760 x 1630	1300 x 760 x 1630	1300 x 760 x 1630	1300 x 760 x 1630	1300 x 760 x 1630

* *Raakaveden ollessa juomavesilaatuista, TDS-arvo maks. 500 mg/l, lämpötila 10 °C ja paine 3 bar.*

BWT Separtec Oy
Puh.: (02) 4367 300
Faksi: (02) 4367 355
Varppeenkatu 28, PL 19
21201 Raisio
info@bwtater.fi
www.bwtwater.fi

Appendix 4. The test run journal

Content of this appendix is confidential and therefore, not published.

Appendix 5. The laboratory analyses of the water samples

Content of this appendix is confidential and therefore, not published.